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16. ABSTRACT The report gives results of an assessment of the impact on the U.S. petroleum refining industry of a possible EPA regulation limiting the level of gaseous refinery sulfur oxide (SOx) emissions. Computer models representing specific refineries in six geographical regions of the U.S. were developed as the basis for determining the impact on the existing refining industry. New refinery construction during the period under analysis (1975-1985) was also considered by development of computer models representing new grassroots refineries. Control of refinery SOx emissions from both existing and new refineries was defined for the purposes of this study by maximum sulfur levels on refinery fuel and on fluid catalytic cracking unit feedstock and by increased sulfur recovery in the Claus plant. The computer models thus constrained were utilized to assess investment and energy requirements to meet the possible regulation and the incremental cost to manufacture all refinery products as a result of the regulation. Parametric studies evaluated the impact of variations in the types of imported crude oils available for future domestic refining and the projected sulfur level of residual fuel oil manufactured in the U.S.

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VOLUME II. DETAILED STUDY RESULTS

by

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APPENDIX A

CRUDE SLATES

The general objective in selecting crude slates for each cluster model was to simulate as closely as possible the average future mixture of crudes which would be run in each refining area represented by the cluster models. Specifically, the crude slates were chosen to simulate as closely as possible the average domestic/foreign mix, sulfur content, API gravity, and other key properties of future crude slates for these refining areas.

A. METHODOLOGY

The basis for forecasting future crude mixes in each cluster model was the Bureau of Mines data on actual crude inputs in 1973 for the six refinery clusters. (See Table A-1.) For future years, the actual mixes in each cluster model were modified in accordance with changes in availability of certain crudes and the addition of major new crude sources (e.g., Alaskan North Slope). The only limitation on the future crude slates was to restrict the number of different crudes used to a manageable level.

The Bureau of Mines data for 1973 shown in Table A-1 indicates the origin of domestic crude inputs by state and foreign crude receipts by country of origin. For the purposes of forecasting model crude slates, certain representative crudes have been chosen to represent crude inputs of a similar quality. Specifically,

- Louisiana crude was used for Louisiana and low-sulfur Texas crudes;
- Oklahoma crude was used for light, sweet crudes produced in the Midcontinent;
- West Texas was used for high sulfur crudes from both Texas and New Mexico;

- Wilmington and Ventura crudes were used for heavy and light Californian crudes, respectively;
- Nigerian Forcados was used for heavy African crudes;
- Algerian Hassi Messaoud was used for light African crudes;
- Arabian Light was considered representative of average Middle East crudes;
- Minas crude represented southeast Asian crude imports; and
- Tia Juana Medium was assumed to typify Venezuelan export crudes.

Table A-2 indicates the actual crude slates assumed for 1973 for each cluster model based on the representative crude methodology just discussed. Table A-2 also includes a comparison of the average sulfur content of the assumed model crude slates given in the table with industry average sulfur content obtained by the EPA from individual company data. As shown, the sulfur contents of the model crude slates check quite closely with actual reported data.

B. MODEL CRUDE SLATES

In projecting crude slates for the cluster models (see Tables A-3 - A-8), specific assumptions have been made regarding crude requirements and availability for each cluster model. These assumptions have been made to simulate the average crude inputs which will prevail in the refining areas represented by the cluster models. The specific crude supply conditions and assumptions for each refining area are as follows:

Small Midcontinent: The cluster model simulating the typical small Midcontinent refiner shows that on average the small Midcontinent refiner currently runs a large proportion of various Texas, Louisiana, and Oklahoma crudes (82%) and small amounts of Canadian (18%). In general, it has been forecast that in the future there will be decreasing amounts of domestic crudes (particularly lower sulfur Louisiana crudes) available to the small Midcontinent refiner and, as a result, he will find himself turning increasingly towards foreign crude sources. (See Table A-3.) It has been assumed that Canadian crude will be rapidly withdrawn from these refiners and that the future foreign crude requirements will be a 50/50 mixture of relatively high sulfur Middle Eastern crudes and low sulfur African crudes. There is already evidence of this

Table A-1. BUREAU OF MINES RECEIPTS OF CRUDE BY ORIGIN 1973 (MB/CD)

Crude source	Louisiana Gulf	Texas Gulf	Small Midcontinent	Large Midwest	West Coast	East Coast
Domestic crudes						
State of orgin:						
Alabama	50	5,231				
Alaska					12,146	
California					89,254	
Colorado			459	3,398		
Florida		12,051				2,594
Illinois			63	9,090		
Kansas			18,906	836		
Louisiana	185,654	18,306	2,259	19,354		3,346
Oklahoma	2,395	1,601	21,931	10,643		
Mississippi						275
Montana				241		
Nebraska			85	836		
New Mexico			370	32,673	678	
Texas	40,502	281,252	4,686	48,805		37,800
Utah					4,321	
Wyoming			1,022	10,818		
Total domestic	228,601	318,441	49,781	136,694	106,399	44,015
Foreign crudes						
Country of origin:						
Algeria						25,248
Angola		3,869				
Canada			10,744	26,022		
Ecuador					7,019	
Indonesia	161					1,165
Iran		3,666			5,970	3,905
Iraq					515	
Libya	214	50				16,121
Mexico		489				
Nigeria	827	10,213		238		29,430
Qatar					2	
Saudi Arabia	910	15,732		4,291	27,056	6,036
Sumatra					24,712	
Trinidad	189					1,772
Tunisia						3,733
United Arab Emirates	546				3,927	5,676
Venezuela	263	7,257			1,295	61,334
Total foreign crudes	3,110	41,276	10,744	30,551	70,446	154,430

Table A-2. ADL MODEL CRUDE SLATES AND SULFUR CONTENTS FOR 1973

(MB/CD)

Crude type	Louisiana Gulf		Texas Gulf		Small Midcontinent		Large Midwest		West Coast		East Coast	
	Volume	%	Volume	%	Volume	%	Volume	%	Volume	%	Volume	%
Domestic												
Louisiana	197.2	88.7	157.2	47.4	4.2	7.6	8.7	6.0	–	–	28.9	15.4
West Texas Sour	25.0	11.3	137.1	41.4	7.2	13.1	102.4	70.3	–	–	14.4	7.6
Oklahoma	–	–	–	–	33.9	61.5	7.2	4.9	–	–	–	–
Calif. Wilmington	–	–	–	–	–	–	–	–	58.0	37.4	–	–
Calif. Ventura	–	–	–	–	–	–	–	–	21.4	13.8	–	–
Subtotal	222.2	100.0	294.3	88.8	45.3	82.2	118.3	81.2	79.4	51.2	43.3	23.0
Foreign												
Nigerian	–	–	12.7	3.8	–	–	–	–	–	–	30.4	16.2
Arabian Light	–	–	17.4	5.3	–	–	12.3	8.5	48.6	31.3	14.2	7.6
Venezuelan	–	–	7.0	2.1	–	–	–	–	–	–	59.6	31.7
Algerian	–	–	–	–	–	–	–	–	–	–	40.5	21.5
Mixed Canadian	–	–	–	–	9.8	17.8	15.0	10.3	11.0	7.1	–	–
Indonesian	–	–	–	–	–	–	–	–	16.2	10.4	–	–
Subtotal	0	0	37.1	11.2	9.8	17.8	27.3	18.8	75.8	48.8	144.7	77.0
Total	222.2	100.0	331.4	100.0	55.1	100.0	145.6	100.0	155.2	100.0	188.0	100.0
Sulfur Content (% weight)												
Model average	0.331		0.765		0.367		1.130		1.251		0.789	
Industry average[a]	0.40		0.72		0.37		1.17		1.30		0.73	

[a] Reference-transmitted to ADL by EPA on 1/22/75

Table A-3. MODEL CRUDE SLATES—SMALL MIDCONTINENT

(MB/CD)

	1977		1980		1985	
	Volume	Percent	Volume	Percent	Volume	Percent
Domestic						
West Texas Sour	6.6	12.0	6.0	10.9	5.5	10.0
Louisiana	3.3	6.0	1.8	2.9	–	–
Oklahoma	32.9	59.8	31.8	57.8	30.7	55.8
Subtotal	42.8	77.8	39.4	71.6	36.2	65.8
Foreign						
Canadian	4.8	8.8	–	–	–	–
Saudi Arabian Light	3.7	6.7	7.8	14.2	9.4	17.1
Algerian Hassi Messaoud	3.7	6.7	7.8	14.2	9.4	17.1
Subtotal	12.2	22.2	15.6	28.4	18.8	34.2
Total	55.0	100.0	55.0	100.0	55.0	100.0

trend towards increasing use of foreign crudes in the two new crude pipelines under construction from the Texas Gulf Coast to Cushing, Oklahoma. Both Texoma and Seaway pipelines were designed to transport foreign crudes to refineries in the Midcontinent.

Large Midwest: Like the small Midcontinent refiners, the large Midwest refiners have typically run a high percentage of domestic crudes (81%) with smaller volumes of Canadian and Middle Eastern crudes. In the future, it is expected that domestic crudes (again, particularly the sweeter Louisiana crudes) will decline in importance and be replaced by foreign crudes (See Table A-4.) Domestic oil policy in Canada will cause Canadian exports to large Midwest refiners to decline rapidly, so that by 1980 no Canadian crudes are likely to be run in Midwest refineries. The demise of Canadian exports to the area will force refiners to rely increasingly on Middle Eastern crudes. Since large Midwest refineries were built to handle high sulfur domestic crudes, it is forecast that higher-sulfur Middle Eastern sources (typified by Saudi Arabian Light) will become increasingly important to the area's refineries. As in the case of the small Midcontinent refineries, the trend towards increased reliance on foreign crude supplies is evidenced by the expansion of the region's only direct, non-Canadian, foreign crude line--Capline--from the Louisiana Gulf.

Texas Gulf: Refiners in the Texas Gulf have run almost 90% domestic crude (48% Louisiana and 42% West Texas Sour). The remaining 10-11% of crude supplies have come from a combination of Middle Eastern, African, and Venezuelan sources. In the future, it is forecast that the crude supply pattern in this region will not change significantly, as decreasing supplies of domestic crudes are reserved for use by refiners in the Gulf area. (See Table A-5.)

East Coast: Historically, typical refineries in this area have run a very high proportion of foreign crude (77% in the cluster model in 1973). As domestic crude supplies decline in the future, it is forecast that East Coast refiners will rely entirely upon foreign crude sources. (See Table A-6.) Because of the severe sulfur restrictions on the East Coast, it is projected that there will be significant imports of Algerian and Nigerian type crudes (representing some 38% of total crude oil to the East Coast

Table A-4. MODEL CRUDE SLATES—LARGE MIDWEST

(MB/CD)

	1977		1980		1985	
	Volume	Percent	Volume	Percent	Volume	Percent
Domestic						
West Texas Sour	93.4	65.1	86.2	60.1	79.0	55.1
Louisiana	4.3	3.0	–	–	–	–
Oklahoma	6.9	4.8	6.6	4.6	6.3	4.4
Subtotal	**104.6**	**72.9**	**92.8**	**64.7**	**85.3**	**59.5**
Foreign						
Canadian	7.4	5.2	–	–	–	–
Saudi Arabian Light	31.4	21.9	50.6	35.3	58.1	40.5
Subtotal	**38.8**	**27.1**	**50.6**	**35.3**	**58.1**	**40.5**
Total	**143.4**	**100.0**	**143.4**	**100.0**	**143.4**	**100.0**

Table A-5. MODEL CRUDE SLATES—TEXAS GULF

(MB/CD)

	1977		1980		1985	
	Volume	Percent	Volume	Percent	Volume	Percent
Domestic						
West Texas Sour	136.0	41.4	136.0	41.4	136.0	41.4
Louisiana	155.7	47.4	155.7	47.4	155.7	47.4
Subtotal	291.7	88.8	291.7	88.8	291.7	88.8
Foreign						
Saudi Arabian Light	17.4	5.3	17.4	5.3	17.4	5.3
Nigerian Forcados	12.5	3.8	12.5	3.8	12.5	3.8
Tia Juana Medium	6.9	2.1	6.9	2.1	6.9	2.1
Subtotal	36.8	11.2	36.8	11.2	36.8	11.2
Total	328.5	100.0	328.5	100.0	328.5	100.0

Table A-6. MODEL CRUDE SLATES—EAST COAST

(MB/CD)

	1977		1980		1985	
	Volume	Percent	Volume	Percent	Volume	Percent
Domestic	–	–	–	–	–	–
Subtotal	–	–	–	–	–	–
Foreign						
Saudi Arabian Light	70.7	35.7	80.5	40.7	90.4	45.7
Algerian Hassi Messaoud	40.8	20.6	38.8	19.6	36.8	18.6
Nigerian Forcados	34.0	17.2	36.0	18.2	38.0	19.2
Tia Juana Medium	52.4	26.5	42.6	21.5	32.7	16.5
Subtotal	197.9	100.0	197.9	100.0	197.9	100.0
Total	197.9	100.0	197.9	100.0	197.9	100.0

cluster model in 1977-1985), with the remainder supplied by Middle Eastern and Venezuelan crudes.

West Coast: The refineries comprising this cluster model currently run an almost 50/50 mixture of domestic Californian crudes and foreign crudes. The predominent domestic crude is presently typified by California Wilmington (a heavy crude). While the West Coast refineries modeled currently consume Middle Eastern, Indonesian, and Canadian crudes, the Middle Eastern crudes account for nearly twice as much as the Canadian and Indonesian inputs combined. In the future it is forecast that volumes of local California crudes will remain at present levels, but foreign crudes will be entirely backed out by Alaskan North Slope volumes beginning in 1980. (See Table A-7 .)

Louisiana Gulf: As simulated in the cluster model, refineries along the Louisiana Gulf presently operate exclusively on domestic crudes. Of these domestic crudes, about 88% are sweet and the remainder sour. Due to the fact that these Louisiana refineries are ideally located to process offshore and southern Louisiana crude production, it is forecast that there will be no change in the crude slate for the typical refiner on the Louisiana Gulf. (See Table A-8 .)

Grassroots Refineries: The crude slates for the grassroots refineries were based on balancing the total crude supply to the U.S.A. on an East of the Rockies and West of the Rockies basis.

East of the Rockies: Grassroots refineries on the East Coast are projected to run a mixture of three types of foreign crudes. Two-thirds of the crude input will be composed of Middle Eastern crudes typified by Saudi Arabian Light; the remaining one-third will be made up of a 50/50 mixture of crudes like Algerian Hassi Messaoud and Nigerian Forcados.

West of the Rockies: Because of the huge volumes of Alaskan North Slope crude expected to be available on the West Coast, it was assumed that grassroots refineries in that part of the country would run exclusively Alaskan North Slope crudes.

C. CRUDE MIX FOR TOTAL U.S.

In order to assess the implications for the total U.S. crude slate of our assumptions regarding crude inputs to cluster models and grassroots

Table A-7. MODEL CRUDE SLATES–WEST COAST

(MB/CD)

	1977		1980		1985	
	Volume	Percent	Volume	Percent	Volume	Percent
Domestic						
California Wilmington	65.7	40.0	65.7	40.0	65.7	40.0
California Ventura	21.7	13.2	21.7	13.2	21.7	13.2
Alaskan North Slope	–	–	76.8	46.8	76.8	46.8
Subtotal	87.4	53.2	164.2	100.0	164.2	100.0
Foreign						
Canadian	5.6	3.4	–	–	–	–
Saudi Arabian Light	54.8	33.4	–	–	–	–
Indonesian Minas	16.4	10.0	–	–	–	–
Subtotal	76.8	46.8	–	–	–	–
Total	164.2	100.0	164.2	100.0	164.2	100.0

Table A-8. MODEL CRUDE SLATES—LOUISIANA GULF

(MB/CD)

	1977		1980		1985	
	Volume	Percent	Volume	Percent	Volume	Percent
Domestic						
Louisiana Sweet	192.3	88.2	192.3	88.2	192.3	88.2
West Texas Sour	25.7	11.8	25.7	11.8	25.7	11.8
Subtotal	218.0	100.0	218.0	100.0	218.0	100.0
Foreign	–	–	–	–	–	–
Subtotal	–	–	–	–	–	–
Total	218.0	100.0	218.0	100.0	218.0	100.0

refineries, we have scaled up the volume of inputs consistent with the amount of refining capacity represented by each cluster or type of grassroots refinery. Table A-9 shows the results of the scale up exercise using the crude oil required in Scenario A. For the cluster models the scaled up crude slates in 1980 and 1985 are constant between scenarios. Crude inputs to the grassroots models were allowed to vary between scenarios to balance required gasoline production. Table A-10 shows the scaled up crude inputs to the grassroots models for 1980 and 1985.

In general, there is expected to be an increasing reliance on foreign crudes in 1980 and 1985 as a result of declines in crude production in the "Lower 48." Even with 1.6 million B/CD of Alaskan North Slope crude consumed in 1985, the decline in production from existing reserves is not offset. Among the foreign imports, the Middle Eastern crudes typified by Saudi Arabian Light are expected to provide by far the largest share, with Nigerian, Algerian, and Venezuelan type crudes accounting for about half of the volume of Middle Eastern crudes.

The scale up crude slate detailed in Table A-9 indicates that the proportions of sweet and sour crude within the total crude slate will not change substantially over the next decade (see also Table A-11). Over the next five years there will be a rise in the volume of sour crudes processed in U.S. refineries as a result of insufficient worldwide production of low sulfur crudes to offset declining U.S. production of sweet crudes. However, this increase in average sulfur content is not expected to cause processing constraints, since domestic and Caribbean downstream processing capacity is forecast to be sufficient to allow refiners to meet sulfur constraints even with a higher average sulfur content in their crude inputs.

In the longer-term, the sweet/sour balance will be preserved despite declines in production from existing domestic sources as a result of an increased availability of sweet crudes on the world market. It is projected that not only will significant volumes of new low-sulfur crude production become available (e.g., the North Sea production and increased Chinese exports), but output from current low-sulfur sources, such as Indonesia will increase. While it is possible that some of the crude from new low-sulfur production sources will be shipped to the U.S., it is

Table A-9. SCALE UP OF MODEL CRUDE SLATES, SCENARIO A

(MB/CD)

Crude type	1980					1985				
	Cluster models	Atypical	Grass Roots	Total		Cluster models	Atypical	Grass Roots	Total	
				Volume	%				Volume	%
Domestic										
Louisiana	3,253.2	–	–	3,253.2	22.1	3,226.2	–	–	3,226.2	19.3
West Texas Sour	3,368.0	–	–	3,368.0	22.9	3,229.0	–	–	3,229.0	19.4
Oklahoma	640.0	–	–	640.0	4.4	616.8	–	–	616.8	3.7
California	1,057.8	–	–	1,057.8	7.2	1,062.3	–	–	1,062.3	6.4
North Slope	930.5	–	396.1	1,326.6	9.0	934.5	–	645.8	1,580.3	9.5
Other	–	698.1	–	698.1	4.8	–	335.1	–	335.1	2.0
Subtotal	9,249.5	698.1	396.1	10,343.7	70.4	9,068.8	335.1	645.8	10,049.7	60.3
Foreign										
Arabian Light	1,850.5	–	852.2	2,703.0	18.4	2,088.0	62.7	1,985.6	4,136.3	24.8
Nigerian	412.3	–	210.1	622.4	4.2	427.4	162.7	489.3	1,079.4	6.5
Algerian	422.2	–	210.1	632.3	4.3	434.1	62.7	489.3	986.1	5.9
Venezuelan	400.2	–	–	400.2	2.7	324.9	100.0	–	424.9	2.5
Canadian	–	–	–	–	–	–	–	–	–	–
Indonesian	–	–	–	–	–	–	–	–	–	–
Other	–	–	–	–	–	–	–	–	–	–
Subtotal	3,085.2	–	1,272.7	4,357.9	29.6	3,274.4	388.1	2,964.2	6,626.7	39.7
Total	12,334.7	698.1	1,668.8	14,701.6	100.0	12,343.2	722.2	3,610.0	16,676.4	100.0

Table A-10. TOTAL CRUDE RUN TO GRASS ROOTS REFINERIES

(MB/CD)

	Scenario					
Year/Grass Roots Region	**A**	**B**	**C**	**D**	**E**	**F**
1980						
West of Rockies	396.1	405.2	412.7	419.9	420.7	
East of Rockies	1,272.7	1,323.7	1,370.8	1,391.4	1,451.5	b
Total	**1,668.8**	**1,728.9**	**1,783.5**	**1,811.3**	**1,872.2**	
1985						
West of Rockies	645.8		672.6	703.3	702.5	687.8
East of Rockies	2,964.2	a	3,192.9	3,240.9	3,506.2	3,240.3
Total	**3,610.0**		**3,865.5**	**3,944.2**	**4,208.7**	**3,928.1**

a. Scenarios B and C are identical in 1985.

b. Scenario F not analyzed in 1980.

Table A-11. DISTRIBUTION OF SWEET AND SOUR CRUDE RUN

(Percent of total crude run)

	1973		1980			1985		
	Sour	Sweet[a]	Sour	Sweet[a]	Other[b]	Sour	Sweet[a]	Other[b]
Domestic								
Louisiana	–	36.1	–	22.1	–	–	19.3	–
West Texas Sour	26.1	–	22.9	–	–	19.4	–	–
Oklahoma	–	3.7	–	4.4	–	–	3.7	–
California	7.3	–	7.2	–	–	6.4	–	–
North Slope	–	–	9.0	–	–	9.5	–	–
Subtotal	33.4	39.8	39.1	26.5	–	35.3	23.0	–
Foreign								
Nigerian Forcados	–	3.9	–	4.2	–	–	6.5	–
Saudi Arabian Light	8.4	–	18.4	–	–	24.8	–	–
Venezuelan	6.0	–	2.7	–	–	2.5	–	–
Algerian Hassi Messaoud	–	3.7	–	4.3	–	–	5.9	–
Canadian	3.3	–	–	–	–	–	–	–
Indonesian	–	1.5	–	–	–	–	–	–
Subtotal	17.7	9.1	21.1	8.5	–	27.3	12.4	–
Other[b]	–	–	–	–	4.8	–	–	2.0
Total	51.1	48.9	60.2	35.0	4.8	62.6	35.4	2.0

[a] "Sweet" crude is defined as crude with a sulfur content less than 0.5%. "Sour" crude includes all crudes with sulfur contents greater than 0.5%.

[b] "Other" refers to crudes which have not been defined by source and whose sulfur content is therefore unknown.

more likely that low-sulfur crudes from new sources will displace low-sulfur material from existing sources in traditional sulfur-sensitive markets. For example, Chinese crude will likely go to Japanese markets where it will displace some of the lower-sulfur Middle Eastern and Indonesian imports. Similarly, North Sea production will be consumed in Europe, freeing up some of the sweet African crudes for delivery to the U.S.

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APPENDIX B

U.S. SUPPLY/DEMAND PROJECTIONS

A. DEMAND ASSUMPTIONS FOR MODEL RUNS

In the current study the demand forecast for the United States was obtained by two different approaches. To ease the process of relating the demand forecast with the scale up of the cluster model approach (Appendix G), one simplistic forecasting approach was utilized which led to a demand growth rate of 2% per annum for all petroleum products. However, to ensure that the study results were not unduly influenced by this simplistic approach, parametric runs were undertaken to include the effect of a more sophisticated forecasting technique. Each of these forecasting techniques is described in detail below.

The primary reason for forecasting demand was to highlight the differences in requirements for grassroots capacity among the six scenarios. Therefore, the actual demand forecast was thought to be of relatively little importance in comparison to the relative differences inherent in the scenarios. Because of the minor importance attached to the absolute projected demand levels, a very simplistic forecasting approach was initially utilized. As will be discussed below, the initial simplistic approach resulted in an overall demand forecast which was well within the range later derived by more elaborate forecasting techniques.

This appendix discusses petroleum product demand only for the U.S. as a whole. To arrive at scaled up product outturns which must be met by the simulated refining industry, imports of petroleum products (assumed to be constant at 1973 actual levels throughout the period of analysis) and outturns from atypical refineries must be subtracted from overall U.S. demand.

From the base year, 1973, product demand was forecast to realize zero growth over 1974 and 1975 and average 2% per annum thereafter. In late 1975, it is evident that demand over 1974-1975 will, indeed, show zero or, in some areas, negative growth relative to 1973 demand. Beyond 1975, public projections of oil demand growth range between 1% and 3.5%, depending upon key assumptions regarding oil prices, consumer price sensitivity, conservation incentives, the availability of alternative energy forms, and U.S. government policy. The estimate of 2% average annual growth was selected as a consensus figure, reflecting the generally slower than historical growth trend which has resulted from higher oil prices, but assuming some optimism regarding the future economic growth of the country.

The methodology initially utilized involved three simplifying assumptions: (1) demand for all products was assumed to grow at one uniform rate (2% per annum from 1975 to 1985); (2) demand growth would occur in equal increments throughout the forecast period; and (3) imports of products would remain constant in volume and type throughout the forecast period.

Table B-1 shows the demand levels for major products which result from an application of the simplified forecasting methodology. It should be noted that these projections were based not on actual 1973 demand, but on a "simulated" demand slate derived by adding the 1973 imports and outturn from atypical refineries to the scaled up yield of each of the cluster models. Since the scale up of 1973 for all products was based on the factors required to have gasoline yields for individual refining areas equivalent to 1973 Bureau of Mines statistics on gasoline outturn in each area, products other than gasoline were not necessarily scaled up to actual consumption figures. The reason for using a "simulated" demand slate as a projection base rather than an actual historical one was that the use of actual data would have resulted in a demand forecast in future years which, by comparison with our projected scaled up cluster outturns, would have accentuated the initial 1973 differences in scaled up cluster outturn and actual demand statistics. Because the differences in projected

Table B-1. PROJECTIONS OF MAJOR PRODUCT DEMAND IN TOTAL U.S. ASSUMED IN MAKING MODEL RUNS

(MB/CD)

	1973 Base year "Simulated Demand"	1973 Base year Actual Demand*	1977	1980	1985
Gasoline	6,706	6,719	6,977	7,404	8,174
Jet fuel & kerosene	1,177	1,275	1,225	1,299	1,432
Distillate fuel oil	3,560	3,092	3,693	3,920	4,328
Residual fuel oil	2,809	2,822	2,922	3,101	3,424
Total-major products	14,271	13,908	14,817	15,724	17,358

*Bureau of Mines.

demand and scaled up cluster outturn are used to set the quality and yield structure of grassroots refineries, it is important that small differences between actual statistics and our scale ups in 1973 not be magnified over the forecast period.

Table B-2 shows a comparison of the forecast of "simulated" demand for major products and the range of demand projections resulting from our more detailed forecast. As indicated, the projection of total "simulated" demand falls within the range of demand forecast by the more elaborate methods which will be described below. However, although there is overall agreement on total demand, there are discrepancies within individual products as a result of: (1) differences in base year starting points (see Table B-1); (2) the assumption of constant average annual growth; and (3) the lack of differential product growth. The differences in base year statistics result from the methods used in scaling up cluster output to regional refinery yields. As described above, one scale up factor (based on the premise of equating gasoline outturns to actual gasoline yields) for all products inevitably resulted in the deviation of some "simulated" product demands from actual reported consumption. The assumption of common growth rate for all products not only reinforced the initial inconsistencies resulting from the scale up procedure but permitted no oscillation in product growth rates. In general, the detailed forecast described below shows product growth (particularly, fuel oils) resuming a moderate growth through the late 1970's and then dropping off in the 1980's as more efficient conservation practices become feasible, more non-oil energy is available and the national economy grows more slowly and with less energy input.

Comparing the total simulated product demand with the demand forecast described below also shows that the results of the simplified approach fall within the range projected in the detailed forecast. Table B-3 compares the total product demands forecast in the simplified and detailed forecasts. Several important assumptions should be noted. Firstly, the LPG demand forecast in the "simulated" demand reflects only the demand for refinery-produced LPG, not LPG extracted at natural gas

Table B-2. A COMPARISON OF PROJECTED "SIMULATED" DEMAND FOR MAJOR PRODUCTS WITH RESULTS OF DETAILED FORECAST

(MB/CD)

	1977			1980			1985		
		Forecast			Forecast			Forecast	
	"Simulated"	High	Low	"Simulated"	High	Low	"Simulated"	High	Low
Gasoline	6,977	7,370	7,190	7,404	8,160	7,520	8,174	9,010	7,520
Jet fuel & kerosene	1,225	1,320	1,300	1,299	1,470	1,390	1,432	1,790	1,600
Distillate fuel oil	3,693	3,420	3,040	3,920	3,870	3,110	4,328	4,270	3,400
Residual fuel oil	2,922	3,140	2,790	3,101	4,030	3,240	3,424	3,020	2,750
Major products total	14,817	15,250	14,320	15,724	17,530	15,260	17,358	18,090	15,270

Table B-3. A COMPARISON OF PROJECTED TOTAL PETROLEUM PRODUCT DEMAND IN "SIMULATED" DEMAND CASE WITH DETAILED FORECAST

(MB/CD)

	1977			1980			1985		
		Forecast			Forecast			Forecast	
	"Simulated"	High	Low	"Simulated"	High	Low	"Simulated"	High	Low
Gasoline	6,977	7,370	7,190	7,404	8,160	7,520	8,174	9,010	7,520
LPG (excl. products at gas processing plants)	513	480	460	545	540	500	603	590	560
Jet fuel & kerosene	1,225	1,320	1,300	1,299	1,470	1,390	1,432	1,790	1,600
Distillate fuel oil	3,693	3,420	3,040	3,920	3,870	3,110	4,328	4,270	3,400
Residual fuel oil	2,922	3,140	2,790	3,101	4,030	3,240	3,424	3,020	2,750
Lubes, waxes & coke	422	450	420	448	490	470	496	540	520
Asphalt	354	530	520	408	590	550	452	660	610
Other	504	1,040	1,010	538	1,100	1,080	597	1,360	1,310
Total	16,610	17,750	16,730	17,663	20,250	17,860	19,506	21,240	18,270
Memo:									
LPG production at gas processing plants		720	680		800	750		890	820
Estimated refinery fuel & losses		1,430	1,410		1,620	1,530		1,780	1,690
Total forecast petroleum consumption		19,900	18,820		22,670	20,140		23,910	20,780

liquids plants. In recent years approximately 65-70% of all LPG originated from natural gas liquids processing, with the remainder supplied from refineries and imports. In the future, the volume of natural gas liquids and the LPG yield from these liquids are expected to decline, so we have assumed in our 1985 projection that only 60% of the LPG will be produced at natural gas processing plants. A second factor to keep in mind in comparing forecasts is that the simplified forecast does not include projections of refinery fuel and losses. Thus, the simplified forecast should not be read as a forecast of total U.S. oil requirements.

B. DETAILED U.S. PRODUCT DEMAND FORECAST

Prior to 1973 forecasting oil demand in the U.S. was a fairly straightforward exercise involving the application of historically-determined growth rates to base year consumption data. However, the pattern of continuous growth was interrupted by massive increases in foreign oil prices (and later domestic decontrolled prices), the Arab oil embargo and a period of economic recession. Oil consumption was largely unaffected in 1973, but beginning in 1974, oil demand actually dropped an estimated 3.8%.

Looking to the future there is a great deal of uncertainty surrounding the level and growth of oil demand. Contrary to historical trends, oil demand is not expected to resume a rapid and steady upward climb. However, the extent to which U.S. demand will turn upward and the timing of the upswing are still very uncertain. In order to bound this uncertainty, Arthur D. Little, Inc., has developed a range of estimates of total energy and oil demand which we feel effectively defines the limits within which future energy demand will probably fall. The methodology underlying the forecast, as well as the forecast itself, is discussed below.

1. Methodology

To forecast future oil consumption it is necessary to evaluate the supplies and demand for all primary energy forms (coal, natural gas, hydroelectric, nuclear and oil). Two basic sets of assumptions were used in order to develop a definitive range of energy supply/demand balances:

High Case: Economic growth is assumed to be somewhat slower than historic rates, but high enough to permit a rising standard of living. Higher energy prices alone--and not government action--are assumed to result in consumer energy conservation. Likewise, higher energy prices are the impetus for the development of domestic energy resources.*

Low Case: Economic and total energy growth fall further off historic rates as a result of both strong government action and higher energy prices. Government action in the form of conservation incentives, selective taxes on oil, import tariffs, etc., is taken to enhance the effects of higher prices in dampening demand and stimulating the development of domestic resources.

The methodology used for deriving oil demand in both cases involved projecting the availability of non-oil energy forms and contrasting these supply estimates with projections of demand consistent with the high and low scenarios. In all cases, oil was assumed to be the balancing fuel in matching the supply and demand estimates. Our forecast of non-oil energy supplies, expressed in quadrillions of Btu's, is given in Table B-4.

Table B-4 shows our projection that total U.S. energy demand will grow at an average of 2.9-3.3% per annum over the period 1972-1980 and between 2.7-4.0% per annum during the first half of the following decade. Oil demand is forecast to grow at between 2.4-3.4% between 1972-1980, and at about 1% in the 1980-85 period.

There are several important features of the primary energy forecast given in Table B-4. Coal production and consumption which have actually declined in recent years are expected to be rejuvenated as a result of higher energy prices, emphasis on exploiting domestic energy resources,

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*This case currently appears to be too optimistic regarding future economic growth and the sensitivity of consumers to rises in energy prices. However, we feel that it does represent the bounding limit on the high side.

Table B-4. PROJECTION OF U.S. PRIMARY ENERGY SUPPLIES WITH OIL AS THE BALANCING FUEL

(Quadrillions of Btu's)

		1980		1985	
	1972	High	Low	High	Low
Coal	12.3	19.6	19.8	23.6	21.5
Natural gas	25.1	23.9	23.9	24.7	24.7
Hydro	2.8	3.2	3.2	3.3	3.2
Nuclear	0.6	5.8	5.8	17.5	14.7
Non-conventional	–	0.1	0.1	0.2	0.2
Oil	32.5	42.5	39.2	44.7	40.9
Total	73.3	95.1	92.0	114.0	105.2

Source: U.S. Bureau of Mines and Arthur D. Little, Inc., estimates.

and production of secondary energy forms from coal. It is expected that it will take the coal industry several years to gear up for higher production levels, but beyond 1980, production capacity will no longer be such a severe limitation on coal consumption. Natural gas is expected to be supply-constrained throughout the forecast period, as production from "Lower 48" resources continues to decline (despite increased offshore activity) and is not offset by volumes from Alaskan sources until very late in the forecast period. Nuclear power is expected to be the most rapidly growing primary energy form, showing a 25-30 fold increase over the forecast period. Non-conventional energy forms, such as solar, wind and solid waste, are not expected to play a significant role during the time frame of this forecast, due to the time required to commercialize and disseminate the technologies involved.

As described above, oil was assumed to be the balancing fuel between the forecast demand for total energy and the projected availability of non-oil energy forms. As such, oil was regarded as being a premium quality fuel, to be increasingly devoted to uses where its liquid, clean-burning properties would command a premium price. Coal, and later nuclear power, were assumed to take over the non-premium oil uses, such as fuel for industrial and utility boilers.

The demand for energy was developed by breaking down total energy consumption into demand by various end-use sectors (e.g., transportation, industry, residential/commercial, etc.). At the end-use sector level the historical growth trends in energy consumption were identified and then modified in line with the basic assumptions of the high and low cases. The modification of historic growth rates took into account our expectations of the impact of consumer conservation, government policy, energy prices, and macro-economic conditions.

The breakdown of oil demand by product, shown in Table B-5, was accomplished by examining the oil consumption patterns of specific end-use sectors. For example, in the transportation sector, the oil demand is a mixture of gasoline, jet fuel, LPG, residual fuel oil, and distillate (diesel). To project future oil consumption patterns in the

Table B-5. FORECAST OF U.S. PRODUCT DEMAND

(MMB/CD)

		1977		1980		1985	
	1975	High	Low	High	Low	High	Low
LPG & refinery gas	1.49	1.68	1.62	1.84	1.75	2.03	1.93
Naphtha & others	0.49	0.56	0.53	0.60	0.58	0.81	0.76
Gasoline	6.56	7.37	7.19	8.16	7.52	9.01	7.52
Kero & jet fuel	1.17	1.32	1.30	1.47	1.39	1.79	1.60
Distillate fuel oil	2.70	3.42	3.04	3.87	3.11	4.27	3.40
Residual fuel oil	2.52	3.14	2.79	4.03	3.24	3.02	2.75
Lubes, waxes, & coke	0.41	0.45	0.42	0.49	0.47	0.54	0.52
Asphalt	0.47	0.53	0.52	0.59	0.55	0.66	0.61
Refinery fuel & losses	1.28	1.43	1.41	1.62	1.53	1.78	1.69
Total	17.09	19.90	18.82	22.67	20.14	23.91	20.78

transportation sector. separate forecasts were developed for automotive, rail, marine and air transport and the fuels were projected accordingly, taking into account any efficiency improvements.

2. Product Forecast

Table B-5 presents our forecast of the range of U.S. demand for petroleum products. This forecast was developed by splitting the total oil demand by end-use sector down into appropriate products and correcting for any changes in consumption patterns.

There are several significant factors to notice about the forecast in Table B-5. Gasoline, which grew at an average annual rate of 5.3% between 1965 and 1972, will grow much more slowly, even in the high demand case. Under the assumptions of the high demand case, gasoline will continue to grow at a modest rate through 1980 (averaging 4.4% per annum), and thereafter growth will be slower (2.0% per annum, 1980-1985) as smaller-engined, more efficient cars penetrate the car population. In the low demand case, we expect gasoline demand to be dampened earlier, with higher gasoline prices, more efficient cars, and, perhaps, government policy combining to keep growth to an average of 2.8% per annum in the 1975-1980 period. In the low demand case, gasoline will actually show no growth over 1980-1985, as the rate of introduction of more efficient cars into the car population offsets the increases in size of the total car population and any increases in average annual mileage travelled.

Jet fuel and kerosene combined are expected to have an average annual growth rate of between 3.5 and 4.8% between 1975 and 1980, and range between 2.9 and 4.0% during the last five years of the forecast period. These forecast growth rates are considerably below the historic growth rates for jet fuel (which averaged just over 6.0% per annum between 1965 and 1972), although kerosene for other uses has historically shown a declining growth trend.

Demand for both distillate and residual fuel oils is expected to grow rapidly until 1980, largely as a result of decreased availability of natural gas and the inability of coal to immediately offset curtailed volumes of natural gas. Beyond 1980, as increased availability of coal and nuclear, and possibly volumes of Alaskan gas, decrease the burden on fuel oils, growth in demand for fuel oil will drop off dramatically. Thus, in the high demand case, demand for distillate fuel oil is projected to grow rapidly at an average of 7.5% per annum between 1975 and 1980, and then decrease dramatically to an average annual rate of 2.0% during the early 1980's. Low case assumptions result in a distillate demand growth of about 3.0% in the 1975-1980 period, declining to 1.8% between 1980 and 1985. Residual fuel oil demand in the high demand case will increase very rapidly through the remainder of the 1970's (averaging just under 10% per annum), and then fall off absolutely in the early 1980's. Similarly, in the low demand case, residual fuel oil consumption is projected to grow at an average annual rate of about 5% between 1975 and 1980, and decrease absolutely at a rapid pace in the 1980-1985 period.

Demand for naphtha and other petrochemical feed stocks is expected to be another area of rapid growth. In the first five years of the forecast, growth in feed stock demand is expected on average to be moderate (averaging between 3.5 and 4.1% per annum) as a result of the recessionary macro-economic conditions in the early years, but rapid growth (between 5.6 and 6.2% per annum) is expected to resume in the early 1980's.

The demand for LPG shown in Table B-5 represents total demand regardless of product source. LPG is presently derived approximately 70% from natural gas processing plants, 22% from refineries, and 8% from imports. In the future, with domestic natural gas production projected to decline (as well as become leaner in natural gas liquids), and Canadian import levels uncertain, there may be pressure on refineries to increase production of LPG, and we anticipate a high level of interest in major LPG import projects.

In summary, total product demand is expected to resume moderate growth through the remainder of the 1970's and then slow to an average of 1% or less in the early 1980's, as conservation (price-induced and/or mandated) and slower rates of national economic growth combine to depress demand below historic levels.

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APPENDIX C

PRODUCT SPECIFICATIONS

Specifications for motor gasolines are presented in Table C-1. Volatility specifications, i.e., Reid vapor pressure and distillation temperatures, were primarily based upon reviewing Bureau of Mines data for summer and winter gasolines. Also, a comprehensive analysis of gasoline volatility is presented in the API study[1] on unleaded gasoline.

In practice, the only distillation specification that exhibited a significant influence on blending flexibility was the maximum percent evaporated at 150°F. Several models, particularly the Texas Gulf, which processed a relatively high percentage of natural gasoline, were impacted by this volatility requirement.

Gasoline distillation end-point specifications are controlled implicitly. For straight-run naphtha the maximum end-point of feed to catalytic reforming is 400°F. For catalytic cracking, the yields and product properties are based on a gasoline/distillate cut point of 400°F in the catalytic cracking fractionation system.

Leaded octane specifications for each cluster model were set identical to the values supplied by the EPA representing actual 1973 operations for the aggregate of the specific refineries comprising each cluster model. The leaded octane requirements for grassroots refining was determined by an average of the cluster models.

The minimum octane requirements for unleaded gasoline were specified at 92/84 research/motor, respectively. Although statutory regulations require only a minimum 91/83 product, indications are that marginally higher octanes are required to provide some allowance for blending tolerances. Specific parametric studies were addressed to manufacturing unleaded grades at higher octane specifications.

Table C-1. PRODUCT SPECIFICATIONS, GASOLINE

Cluster model	Maximum Reid vapor pressure	% Evaporated at 150°F Minimum	% Evaporated at 150°F Maximum	% Evaporated at 210°F Minimum	% Evaporated at 210°F Maximum	Minimum leaded research octane number	Minimum leaded motor octane number
Premium							
East Coast	10.5	20.0	28.0	42.0	54.0	100.5	92.5
Large Midwest	10.5	20.0	30.0	42.0	54.0	99.2	92.6
Small Midcontinent	10.5	20.0	28.0	42.0	54.0	98.9	94.0
Louisiana Gulf	10.5					99.8	92.2
Texas Gulf	10.5	20.0	28.0	42.0	54.0	99.5	92.0
West Coast	10.5	20.0	28.0	42.0	54.0	99.3	90.2
East of Rockies Grassroots	10.5	20.0	28.0	42.0	54.0	99.8	92.7
West of Rockies Grassroots	10.5	20.0	28.0	42.0	54.0	99.3	90.2
Regular							
East Coast	10.5	20.0	30.0	42.0	54.0	94.1	86.1
Large Midwest	10.5	20.0	30.0	42.0	54.0	94.0	86.2
Small Midcontinent	10.5	20.0	30.0	42.0	54.0	92.2	86.1
Louisiana Gulf	10.5					93.6	86.6
Texas Gulf	10.5	20.0	30.0	42.0	54.0	93.8	86.8
West Coast	10.5	20.0	30.0	42.0	54.0	93.4	84.5
East of Rockies Grassroots	10.5	20.0	30.0	42.0	54.0	93.9	86.4
West of Rockies Grassroots	10.5	20.0	30.0	42.0	54.0	93.4	84.5
Unleaded							
Louisiana Gulf	10.5					92.0	84.0
Texas Gulf	10.5	20.0	32.0	42.0	54.0	92.0	84.0
All Others	10.5	20.0	30.0	42.0	54.0	92.0	84.0

Other product specifications are presented in Table C-2. The basic philosophy was to adopt only those key specifications necessary to measure the impacts being evaluated in this study. For example, there are approximately 20 different specifications on commercial jet fuel, each of which is critical to satisfactory performance. Yet only several key ones are required for use in an aggregate model of this type. It is only meaningful to consider product specifications for which specific processing adjustments are required which affect unit costs. Those specifications that are met by segregated blending, while they require careful planning in the individual refineries, are not relevant to this level of simulation.

For example, the smoke-point specification for jet fuel must be met on all products supplied. However, in the East of Rockies system, there are ample supplies of good quality blend stocks such that no special processing or additives are needed. On the other hand, in the West Coast one of the reasons for the widespread installation of hydrocracking is due to the relatively high product demands for jet fuel and the relatively poor smoke-point of product produced from indigenous crudes. Thus, we do specify that the smoke-point requirement be met by the model for the West Coast. Although no sulfur specification was used in the model for jet fuel, the model is structured such that only desulfurized components can be routed to jet fuel blending with the exception of the Louisiana cluster model which processes low sulfur crudes.

The initial end-point distillation requirements for jet fuel (as well as for other products) are met by the model structural control of fractionating cut points, which allow only "specification" components to be made available for blending.

Distillate fuel oil sulfur specifications vary from cluster model to cluster model. They were adjusted during the calibration phase to achieve a reasonable utilization of existing desulfurization facilities and the specifications thus determined were then used for the simulation at future years. Diesel fuels are included in the general distillate fuel category. Although they require various cetane number specifications in the marketplace, they are met by blending and thus need not be considered in this analysis.

Table C-2. OTHER PRODUCT SPECIFICATIONS

Cluster Model	Minimum specific gravity	Maximum sulfur level – % wt	Minimum smoke point – mm	Viscosity – Refutas @ 122°F Minimum	Viscosity – Refutas @ 122°F Maximum
Jet fuel					
West Coast and West of Rockies Grass Roots	0.797		20.0		
All others	0.797				
Kerosene					
All clusters	0.797	0.1			
Distillate fuel oil					
East Coast		0.1			
Large Midwest		0.2			
Small Midcontinent		0.2			
Louisiana Gulf		0.1			
Texas Gulf		0.17			
West Coast		0.14			
East of Rockies Grass Roots		0.1			
West of Rockies Grass Roots		0.1			
Residual fuel oil					
East Coast		2.0		28.0	37.0
Large Midwest		1.5		28.0	37.0
Small Midcontinent		1.5		28.0	37.0
Louisiana Gulf		1.5			
Texas Gulf		1.5			37.0
West Coast		1.0		24.0	37.0
East of Rockies Grass Roots					
Scenario A		1.78		28.0	37.0
Scenario C		1.97		28.0	37.0
Scenario D		2.45		28.0	37.0
Scenario E		2.40		28.0	37.0
Scenario F		2.12		28.0	37.0
West of Rockies Grass Roots					
Scenario A		1.47		28.0	37.0
Scenario C		1.63		28.0	37.0
Scenario D		1.38		28.0	37.0
Scenario E		1.38		28.0	37.0
Scenario F		1.16		28.0	37.0

A similar philosophy was used for residual fuel oil, the critical specifications for which are presented in Table C-2. The variability in sulfur content reflects the sulfur content of products from each geographical area,[2,3] as discussed in Volume I, Section II.

The sulfur content of the refinery fuel oil system is discussed in detail in Appendix D.

References

1. *U.S. Motor Gasoline Economics, Volume 1, Manufacture of Unleaded Gasoline*, American Petroleum Institute, 1967.

2. U.S. Dept. of Interior, Bureau of Mines, Petroleum Products Survey, Burner Fuel Oils (1974).

3. U.S. Dept. of Interior, Bureau of Mines, "Availability of Heavy Fuel Oils by Sulfur Level", December (1973).

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APPENDIX D

BASE LEVEL OF CLUSTER REFINERY FUEL SULFUR CONTENT

U.S. refinery fuel sulfur oxide (SO_x) emission levels are currently controlled primarily by state regulations on fuel sulfur content or on ground level concentrations of sulfur dioxide (SO_2). Since the control of refinery fuel SO_x emissions will place an important constraint on the modeling of the refinery operations, the allowable refinery fuel sulfur content under current or immediately anticipated regulations had to be determined for each cluster model. This is complicated by the regional modeling approach since emission regulations are typically established on a state-by-state basis. This appendix describes the methodology used in calculating emission regulations for each cluster model, beginning with the determination of state regulations and then translating these into equivalent cluster regulations.

Specific regulations on sulfur content of fuels were promulgated not as a means of controling fuel sulfur content itself but as one of the control techniques for the ambient air quality standards on SO_x. Thus, not all states have established explicit standards on fuel sulfur content. While the Federal government has promulgated the National Ambient Air Quality Standards (NAAQS) for SO_x, these regulations represent minimum standards to be achieved. As a result, air quality regulations and, indirectly, fuel sulfur regulations, vary from state to state and often within a state. Since actual implementation of NAAQS is the responsibility of state governments, the means by which regulations are expressed also varies between and within states. For example, some states regulate SO_x emissions by determining the maximum allowable fuel sulfur content for achievement of the particular SO_x standards, while other states express standards only in terms of ambient air or ground level SO_2 concentrations. In the latter case, allowable SO_2 concentrations at the stack have to be determined by dispersion models reflecting local meteorological conditions. For consistent use in the ADL refinery model, this stack concentration was then translated into an equivalent refinery fuel sulfur content, and, in some cases, it was necessary to contact refiners or EPA regional offices where these calculations were available.

A final complicating factor is that these regulations also vary according to the size of the combustion units. This made it necessary to identify the location of refineries within each state and to make simplifying assumptions about the size of refinery combustion units.

A. METHODOLOGY OF CALCULATIONS

1. Fuel Oil Sulfur Content by State

The National Summary of State Implementation Plan Reviews (Volume II), published in July, 1975, by the EPA, was used as the primary source of state regulations on fuel sulfur levels. For those states with fuel sulfur regulations varying by Air Quality Control Region (AQCR), each refinery in that state was identified as to its location in an AQCR. The state regulation was then calculated as the average sulfur regulation of the AQCR's containing refineries, weighted by 1973 crude refining capacity of each AQCR.

Reasonable assumptions were required for those states that did not have explicit fuel oil sulfur regulations. In Texas, the ground level SO_2 concentration regulation may require a maximum liquid fuel sulfur content of 0.7-0.9% wt. for compliance, according to one refiner's modeling of operations. Similarly, in California, the ground level SO_2 concentration translates into fuel sulfur limits of 0.5% wt. Although Louisiana has promulgated a maximum fuel sulfur level of 3.6% wt., ground level concentration standards are, in fact, controling and refiners in some AQCR's limit fuel burned to 0.7-0.9% wt. S. Current regulations in Ohio and Illinois are uncertain at present and, after consultation with local AQCR authorities, were assumed to be 1.0% wt. S.

2. Combustion Unit Size

State regulations for existing and new sources are reported in the above publication by size of combustion unit (heat inputs of 10, 100, 250, or 1000 million Btu/hour), by type of fuel burned, and by air quality control region. Regulations applicable to refineries were taken as those pertaining to residual fuel-fired units with heat inputs of less than 1000 million Btu/hour. Where they differed among the 10, 100 and 250 million Btu/hour units, an arithmetic average of the three was used. If the regulations were explicit only for a 1000 million Btu/hour unit, this was assumed to apply.[1]

B. RESULTS

Table D-1 shows existing and new source fuel sulfur regulations by state which were calculated using the above methodology.

Refinery fuel sulfur level standards by Petroleum Administration for Defense (PAD) districts were then calculated as the weighted average of state regulations comprising each PAD, with crude throughput of each state used as weighting factors. Refinery crude throughput by state is available in the Bureau of Mines 1973 Annual Petroleum Statement. However, for some PAD districts, crude throughput for several states is reported together; where this occurred a weighted average sulfur standard for those states was determined using 1973 state refining capacities as weighting factors.

Table D-2 shows calculated PAD regulations for existing and new sources. Also shown is the state within each PAD with the most and least stringent sulfur regulation. Those states or portions of states for which sulfur standards could not be determined (Western Pennsylvania, Missouri, Oklahoma for existing sources, and Arkansas) were eliminated from the determination of weighted average PAD regulations.

As discussed in Appendix F, specific clusters of refineries were selected to represent typical refineries in each PAD district, to be scaled up to represent the U.S. petroleum industry. For comparison to the average

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[1] Sulfur standards applying to 1000 million Btu/hr heat input units are more applicable to large steam generating units such as electric utilities. In addition, Federal New Source Performance Standards (NSPS) technically apply only to fossil-fuel fired steam generators with heat inputs greater than 250 million Btu/hr. The occasional use of these standards for determining regulations pertaining to refineries was necessitated by lack of suitable alternative information, but it is felt the resulting cluster regulation is nevertheless representative.

Table D-1. REFINERY FUEL SULFUR REGULATIONS BY STATE

PAD	State	1973 Refinery crude throughput (MB/D)	Calculated regulation – % wt. S	
			Existing source	New source
I	New Jersey	596.885	0.3	0.3
	Delaware, Maryland	128.301	0.93	0.93
	Virginia, Georgia, Florida	57.764	2.45	2.44
	New York	100.496	2.2	2.2
	Pennsylvania-East	543.641	0.3	0.3
	West Virginia	13.718	2.7	2.0
	Total	1,440.805		
II	Ohio	500.315	1.0	1.0
	Indiana	491.614	3.63	3.63
	Illinois	1,031.118	1.0	1.0
	Kentucky, Tennessee	175.148	2.05	1.60
	Michigan	122.885	1.16	1.16
	Minnesota, Wisconsin	193.252	1.83	1.83
	N. Dakota	49.101	2.7	2.7
	Oklahoma	447.162	b	0.3
	Kansas	373.266	2.8	0.7
	Total	3,383.861		
III	Texas	3,209.112	0.9	0.7
	Louisiana	1,462.088	0.9	0.7
	Mississippi	256.033	2.2	2.2
	Alabama	31.559	2.54	2.54
	New Mexico	46.389	0.9	0.9
	Total	5,005.181		
V	California	1,577.197	0.5	0.5
	Other States[a]	398.173	1.88	1.87
	Total	1,975.370		
a. PAD V "Other States:"				
	Washington		2.14	2.14
	Oregon		1.4	1.4
	Arizona		0.9	0.7
	Alaska		0.5	0.5
	Hawaii		2.0	2.0

b. Various maximum ambient concentration limits.

Table D-2. REFINERY FUEL SULFUR REGULATIONS BY PAD

PAD	Existing sources – % wt. S			New sources – % wt. S		
	Most restrictive – (state)	Least restrictive – (state)	PAD weighted average	Most restrictive – (state)	Least restrictive – (state)	PAD weighted average
I	0.3 (New Jersey) (Pennsylvania)	2.83 (Georgia)	0.60	0.3 (New Jersey) (Pennsylvania)	2.83 (Georgia)	0.59
II	1.0 (Ohio) (Illinois)	3.63 (Indiana)	1.82	0.3 (Oklahoma)	3.63 (Indiana)	1.37
III	0.9 (Texas) (Louisiana) (New Mexico)	2.54 (Alabama)	0.98	0.7 (Texas) (Louisiana)	2.54 (Alabama)	0.79
V	0.5 (California) (Alaska)	2.14 (Washington)	0.78	0.5 (California) (Alaska)	2.14 (Washington)	0.78

PAD fuel sulfur regulations of Table D-2, the regulations applicable to each of the individual refineries in the clusters are shown in Table D-3. Because of the AQCR's in which these particular refineries are located, the regulations for these refineries are generally more restrictive than the average regulations for the PAD district as a whole.

C. CLUSTER MODEL REFINERY FUEL SPECIFICATION

From Tables D-1, D-2 and D-3, the existing fuel sulfur regulations (to be applied to the sulfur level of the refinery fuel system) for the cluster models must be selected. As noted above, however, there is a conflict between the regulations pertaining to the specific refineries simulated in the cluster models and the average of the PAD district. This makes the selection of represantative refinery fuel sulfur regulations difficult.

On the one hand, if the stringent regulations typical of the specific cluster refineries (Table D-3) are selected to represent the existing maximum refinery fuel sulfur levels in the cluster models, then the computer simulation would indicate that little additional investment would be required to meet future regulations. Since the average PAD district fuel sulfur regulation (Table D-2) is much higher than these most stringent regulations, it is likely that the cost of future sulfur regulations determined by such computer models would understate the actual costs incurred by the industry.

On the other hand, if existing maximum sulfur levels of the refinery fuel system were selected equal to the average PAD district regulations (Table D-2), then the sulfur balances in the cluster models will not match those of the actual refineries being simulated (Appendix F), for these refineries are operated to match the regulations of Table D-3. For some clusters, such as the Texas Gulf Coast, there was such a small fraction of residual oil in the fuel system in 1973 that this inconsistency is unimportant. For the East Coast cluster, however, the inconsistency is significant (compare Table D-2 and D-3).

Since it is felt that understatement of the cost of future regulations is a far more serious error when the industry model is to provide input to environmental policy decisions, the current cluster model fuel regulations

Table D-3. REFINERY FUEL SULFUR REGULATIONS APPLICABLE TO INDIVIDUAL REFINERIES IN CLUSTER MODELS

Cluster	Refinery and location	Fuel regulations – % wt. S Existing source	 New source
East Coast	ARCO Philadelphia, Pa.	0.3	0.3
	Sun-Marcus Hook, Pa.	0.3	0.3
	Exxon-Linden, N.J.	0.3	0.3
Small Midcontinent	Skelly-El Dorado, Kansas	2.8	0.7
	Gulf-Toledo, Ohio	1.0	1.0
	Champlin-Enid, Oklahoma	a	0.3
Large Midwest	Mobil-Joliet, Ill.	1.0	1.0
	Union-Lemont, Ill.	1.0	1.0
	ARCO-E. Chicago, Ill.	1.0	1.0
Texas Gulf	Exxon-Baytown, Texas	0.9	0.7
	Gulf-Pt. Arthur, Texas	0.9	0.7
	Mobil-Beaumont, Texas	0.9	0.7
Louisiana Gulf	Gulf-Alliance, La.	0.9	0.7
	Shell-Norco, La.	0.9	0.7
	Citgo-Lake Charles, La.	0.9	0.7
West Coast	Mobil-Torrance, Ca.	0.5	0.5
	Arco-Carson, Ca.	0.5	0.5
	Socal-El Segundo, Ca.	0.5	0.5

a. Various maximum ambient concentration limits.

were assumed to be those of the PAD district and not of the refineries simulated by the cluster model. The allowable refinery fuel sulfur content assumed in the cluster models is summarized in Table D-4. The East Coast cluster entry is taken as the PAD I regulation of Table D-2. The Small Midcontinent and Large Midwest entries were taken as the average PAD II regulation for existing and new sources of Table D-2 (recall the uncertainty for the Ohio and Illinois entries of Table D-1, making more precise assessments unwarranted). The Texas and Louisiana entries were taken directly from Table D-1, and agree well with Table D-2. The West Coast cluster entry is near that of Table D-2, weighted downward by the California entry of Table D-1.

Furthermore, since most existing refineries must meet these regulations on a stack-by-stack basis (not averaged over total gaseous and liquid fuel systems), these sulfur specifications were used to limit the sulfur level of the liquid fuel system (e.g., residual fuel oil) in the existing cluster refineries. For the grassroots refineries, which would likely have one large stack, these sulfur specifications were applied to the average sulfur levels of liquid and gaseous fuels.

Table D-4. BASE LEVEL OF CLUSTER REFINERY FUEL SULFUR CONTENT USED IN MODEL RUNS

Refinery cluster	Maximum allowable sulfur level in refinery fuel system, wt. %
East Coast	0.6
Small Midcontinent	1.5
Large Midwest	1.5
Texas Gulf	0.9
Louisiana Gulf	0.9
West Coast	0.7

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APPENDIX E

CAPITAL INVESTMENT FOR PROCESS UNIT SEVERITY UPGRADING AND UTILIZATION OF CAPACITY ALREADY CONSTRUCTED

In order to meet the environmental regulations under study, it is often required to utilize existing process unit capacity or to upgrade the process unit severity beyond that required without the proposed regulation. Hence, in the evaluation of investment penalties associated with the regulations, costs must be assessed for (1) the value of the existing facilities which have been utilized and (2) the added expense of upgrading these facilities to meet these regulations.

For example, for the lead regulations, existing catalytic reformer capacity must be utilized to make increasing amounts of unleaded and low lead gasolines. This capacity could be otherwise utilized to produce increasing quantities of leaded gasoline were it not for the lead regulations. Hence, a value must be placed on this excess reformer capacity, and a cost assessed due to lead removal if the capacity utilization exceeds the case without the lead regulations. This cost assessment for existing facilities is well-known in economic theory, and the value of such facilities is obtained from an evaluation of their "opportunity cost" or "alternative value".

The production of unleaded gasoline will require that existing reformers be operated at 100 RON severity, even though the existing reformers may have been designed to operate only at 90 RON severity. Hence, a capital outlay may be required to upgrade the severity capability of such reformers to 100 RON.

The present appendix provides the methodology used in assessing the capital investment penalties for existing capacity utilization and severity upgrading, as well as the results of this calculation for the example clusters of existing refineries. Additional costs of promulgated lead regulations and other possible regulations are incurred in new grassroots refineries, which are summarized in Appendices H and J. Finally, the investments required in these individual clusters are multiplied by scale up factors (Appendix G) to reflect the industry-wide costs, the results of which are tabulated in Appendix J.

From an examination of the unit throughput/severity results of Appendix J, only four refinery units vary sufficiently to warrant investment evaluations for capacity utilization and severity upgrading. These are catalytic reforming, hydrocracking, isomerization and alkylation. Other units, such as fluid catalytic cracking, did not undergo sufficient change in unit intake or severity between scenarios to justify such calculations.

A. CATALYTIC REFORMING

Calibration runs for each of the clusters were performed, as discussed in Appendix I, comparing the reformer throughput required by the cluster model with that obtained from industry data. Also summarized in Appendix I was the comparison of these figures with the stream-day capacity reported in the Oil and Gas Journal.

In Table E-1 are summarized the results of this comparison for each cluster. The model output provided the catalytic reformer intake of high severity (>95 RON) and low severity (< 95 RON) operation required to meet pool octane for each cluster (Appendix I). It also provided the BTX reformer throughout required to meet the 1973 demand for BTX from each cluster.

The existing reformer calendar day capacity in Table E-1 was obtained by multiplying the Oil and Gas Journal stream day capacity (Appendix I) by 85%. This figure represents limitations such as:

- Scheduled or extraordinary refinery turnarounds and maintenance.
- Limitations on secondary processing capacity which can limit meeting product specifications.
- Variations in crude slates since nominal capacity is based on a design crude and would be higher or lower depending on the gravity

of the actual crude run.

- Forced outages due to fires or strikes.
- Crude supply restrictions, particularly those refineries tied to local crudes.
- Regional and logistical constraints.
- Imbalances between individual product output and market demand.

Discussions with industry sources indicated that about three-fourths of total reforming capacity is capable of only low severity operation (high pressure units). The BTX capacity is conservatively taken to be equal to that of the 1973 calibration run, with the remaining capacity available for motor gasoline production. Hence, for the East Coast, Large Midwest and Louisiana Gulf clusters, the existing low severity capacity available was calculated as three-fourths of total reforming capacity, and total high severity capacity was obtained by difference. High severity capacity available for gasoline production in these three clusters was determined as total high severity capacity less 1973 BTX capacity. Applying this same methodology to the Small Midcontinent, Texas Gulf, and West Coast clusters would result in total high severity capacity less than that required for BTX production. Therefore, for these clusters it was assumed that no high severity capacity existed for gasoline production. If total calibration throughput requirements exceeded the adjusted Oil and Gas Journal capacity, the former figure was taken as the capacity limitation.

It is recognized that the zero high severity capacity for motor gasoline production indicated in Table E-1 is not true in reality. However, the assumptions about the amount of existing high and low severity capacity affect only the assignment of penalties to severity upgrading versus utilization of existing capacity. Total capital requirements are not affected, so no further refinement of high severity capacity availability was undertaken.

Next, the cost of upgrading the catalytic reformer severity and the value of existing facilities had to be assessed. After discussions with process licensors, it was determined that the cost of upgrading a catalytic reformer to be capable of 100 RON operation (furnace costs and pressure

Table E-1. CATALYTIC REFORMING CAPACITY AVAILABILITY

(MB/CD)

	Cluster model					
	East Coast	Large Midwest	Small Midcont	Louisiana Gulf	Texas Gulf	West Coast
1973 Calibration throughput						
High severity mogas	7.0	0	0	0	0	0
Low severity mogas	29.0	25.3	10.2	28.3	50.4	21.0
BTX	3.5	2.3	4.3	0	20.3	16.3
Spare capacity	2.4	0.2	0	6.1	0	0
Subtotal	**41.9**	**27.8**	**14.5**	**34.4**	**70.7**	**37.3**
Existing capacity, 1973						
High severity mogas	7.0	4.6	0	8.6	0	0
Low severity mogas	31.4	20.9	8.9	25.8	49.7	20.5
BTX	3.5	2.3	4.3	0	20.3	16.3
Subtotal	**41.9**	**27.8**	**13.2**	**34.4**	**70.0**	**36.8**

alteration) would approximate that of the initial investment of the low severity unit. This investment will certainly vary from site-to-site; hence, this upgrading cost is assumed to be equivalent, but all investment penalties will be reported with this item explicitly identified so that it may be adjusted easily if improved cost data becomes available. Note, however, that only 75% of the units potentially need upgrading.

The value of the existing facilities, to be used in the assessment of the cost of utilization of existing capacity, is more difficult to identify, since it depends upon the value of alternate use of these facilities. Fortunately, the total investment penalty is not heavily dependent upon the value placed on these existing facilities. If the value is taken equal to cost of installation of a low severity unit, the cost of utilization is only 7% of the severity upgrading cost for catalytic reforming.

The investment penalties for each of these items is summarized in Table E-2, along with the cost of installation of new, low-pressure reformers for comparison.

An example calculation of the cost of capacity utilization for the Large Midwest cluster is shown in Table E-3, for Scenarios A and C. The reformer throughput required to meet the gasoline demand and the quality specifications is obtained from the computer output (Appendix J). For Scenario A, few changes take place between 1977 and 1985; for Scenario C, the amount of high-severity reforming increases due to lead regulations, and the total reforming capacity increases due to yield losses. As noted in Table E-3, subtracting the throughput of the prior time period from the throughput requirement allows a break-down between spare capacity utilization and new reformer capacity requirements. After adjusting the stream day investments of Table E-2 to a calendar day basis, the total capacity-related costs can be determined. Note that the cost of spare capacity utilitization is quite low, making the precise "opportunity cost" assessed for this utilization immaterial.

The cumulative onsite cost for Scenario C is $3.37 million for this cluster, while that for Scenario A is $1.40 million. The penalty for lead regulations is thus $1.97 million. This penalty was further increased by

Table E-2. CATALYTIC REFORMER INVESTMENT FOR CAPACITY UTILIZATION AND SEVERITY UPGRADING

	Small Midcontinent cluster		All other clusters	
	Standard unit, MB/SD	Investment, $/B/SD	Standard unit, MB/SD	Investment, $/B/SD
Existing capacity value	10	1,135	25	655
Severity upgrading cost	10	1,135	25	655
New capacity cost	10	1,450	25	780

Table E-3. COSTS OF ADDITIONAL REFORMER CAPACITY

Large Midwest Cluster

	Scenario A			Scenario C		
	1977	1980	1985	1977	1980	1985
Reformer throughput required (MB/CD)						
High severity	4.4	3.7	3.9	17.7	26.7	29.1
Low severity	20.7	20.8	23.0	9.6	0	0
Subtotal	25.1	24.5	26.9	27.3	26.7	29.1
Less existing thruput (MB/CD)	25.3[a]	25.3[a]	25.3[a]	25.3[a]	27.3[b]	27.3[b]
Subtotal (MB/CD)	0	0	1.6	2.0	0	1.8
Less spare capacity (MB/CD)	–	–	0.2	0.2	–	0
New capacity required (MB/CD)	0	0	1.4	1.8	0	1.8
Cost of spare capacity[c] ($MM)	0	0	0.15	0.15	0	0
Cost of new capacity[d] ($MM)	0	0	1.25	1.61	0	1.61
Total capacity cost ($MM)	0	0	1.40	1.76	0	1.61
Cumulative total capacity cost ($MM)	0	0	1.40	1.76	1.76	3.37

[a] 1973 calibration thruput, Table E-1

[b] Total capacity available after 1977

[c] Thruput x (655/0.85)

[d] Thruput x (760/0.85)

40% to allow for offsite costs and working capital, leading to $2.76 million. It should be noted that the investment penalties indicated in Appendix J of "The Impact of Lead Additive Regulations On The Petroleum Refining Industry, Volume II" were calculated on a Scenario B versus Scenario A, and Scenario C versus Scenario B basis. The Scenario C versus A comparison in this appendix is intended only for illustrative purposes. Furthermore, in Appendix J, investments for individual units are reported for onsite costs only, prior to adjustment to a calendar-day basis. The sum of unit onsite costs is first increased by 40% to allow for offsite costs and working capital and then adjusted for stream day investments.

An example calculation of the cost of severity upgrading for the Large Midwest cluster is shown in Table E-4 for Scenarios A and C. The high severity reformer throughput required to meet gasoline demand and quality specifications is identical to that shown in Table E-3. The high severity capacity available was reported in Table E-1 for 1973. Subtraction of the capacity from the throughput requirements and allowing for new high severity capacity constructed from Table E-3 provides the amount of severity upgrading required. After correcting the stream day investments of Table E-2 to a calendar day basis, the cost of severity upgrading can be determined.

The cumulative on-site cost for severity upgrading in Scenario C becomes $16.11 million for the Large Midwest cluster. Subtracting the upgrading cost for Scenario A (zero in this case), the penalty for lead regulations becomes $16.1 million. Increasing this cost by 40% to allow for offsites and working capital, the total penalty for severity upgrading becomes $22.6 million.

Combining the results of Tables E-3 and E-4, the total reformer related capital investment penalty for Scenario C versus Scenario A becomes $25.4 million. To obtain the contribution of the Large Midwest cluster to the U.S. refining industry requires further multiplication of this penalty by the scale-up factor of Appendix G.

B. HYDROCRACKING

The methodology used to assess investment costs for utilization of spare capacity and alterations in severity of operations for hydrocracking is similar to that used for reforming. Again, calibration results indicated

Table E-4. COST OF SEVERITY UPGRADING

Large Midwest Cluster

	Scenario A			Scenario C		
	1977	1980	1985	1977	1980	1985
High severity reformer thruput required (MB/CD)	4.4	3.7	3.9	17.7	26.7	29.1
Less high severity capacity available (MB/CD)	4.6[a]	4.6[a]	4.6[a]	4.6[a]	17.7[b]	26.7[c]
Subtotal	0	0	0	13.1	9.0	2.4
Less new capacity construction (MB/CD)	0	0	0	1.8[d]	0	1.8[e]
Severity upgrading required (MB/CD)	0	0	0	11.3	9.0	0.6
Cost of severity upgrading[f] ($MM)	0	0	0	8.71	6.94	0.46
Cumulative cost of severity upgrading ($MM)	0	0	0	8.71	15.65	16.11

[a] 1973 Existing capacity (Table E-1)
[b] Total capacity available after 1977
[c] Total capacity available after 1980
[d] New capacity built in 1977 (Table E-3)
[e] New capacity built in 1985 (Table E-3)
[f] Thruput x ($655/0.85)

the hydrocracking capacity and severity required to meet 1973 product demands and specifications. Table E-5 shows the results taken from computer outputs of calibration runs. All clusters except the Large Midwest and Small Midcontinent contained hydrocracker units in 1973. For the Louisiana Gulf and West Coast clusters, calibration hydrocracker utilization was at medium severity only, while the Texas Gulf cluster ran at high severity and the East Coast cluster required both high and medium severity operations. The existing capacity shown in Table E-5 is the Oil and Gas Journal stream day capacity (Appendix I) multiplied by 85%. Because published information does not provide a breakdown of existing hydrocracking capacity by level of severity, it was assumed for purposes of calculating investment penalties that actual 1973 hydrocracking severities paralleled the results obtained in calibration. Hence, for example, the existing 18.1 MB/CD hydrocracking capacity for the Texas Gulf cluster was assumed to be designed for high severity operations since calibration results indicated a requirement of 14.7 MB/CD high severity hydrocracking in order to meet the 1973 product demand slate. Table E-5 also shows spare capacity available as the difference between total existing capacity in 1973 and total calibration throughput.

As discussed in the preceeding section on catalytic reforming, an implicit opportunity cost for utilization of spare capacity specifically for meeting promulgated or potential EPA regulations must be incorporated in addition to explicit investment costs for new capacity and for severity changes. The charge for existing capacity has been taken as the investment cost required for new capacity. The cost of changing severity levels, which is 20% of new grassroots investment, represents alterations to provide the flexibility to vary severity of operations as required to meet product demand. Table E-6 shows the investment penalties associated with utilization of spare capacity (i.e., investment costs for new hydrocracking capacity of high and medium severity) and severity flexibility.

A calculation of investment penalties for capacity utilization in the Texas Gulf cluster for Scenarios A and C is provided in Table E-7. In this example, the cost for utilization of existing capacity is the $1270/B stream-day investment for high severity hydrocracking (Table E-6), adjusted to a calendar day basis, since all existing capacity in the Texas Gulf

Table E-5. HYDROCRACKING CAPACITY AVAILABILITY

(MB/CD)

	Cluster model					
	East Coast	Large Midwest	Small Midcontinent	Louisiana Gulf	Texas Gulf	West Coast
1973 Calibration throughput						
High severity	6.2	–	–	–	14.7	–
Medium severity	1.3	–	–	6.6	–	22.1
Total	7.5	–	–	6.6	14.7	22.1
Existing capacity, 1973						
High severity	7.2	–	–	–	18.1	–
Medium severity	1.3	–	–	8.1	–	23.6
Total	8.5	–	–	8.1	18.1	23.6
Spare capacity available	1.0	–	–	1.5	3.4	1.5

Table E-6. HYDROCRACKING INVESTMENT FOR CAPACITY UTILIZATION, NEW CAPACITY, AND SEVERITY FLEXIBILITY

	Standard unit, MB/SD	Investment, $/B/SD
Existing and new capacity cost		
– High severity	30	1270
– Medium severity	30	1090
Cost of severity flexibility	30	240

Table E-7. COSTS OF ADDITIONAL HYDROCRACKING CAPACITY

Texas Gulf Cluster

	Scenario A			Scenario C		
	1977	1980	1985	1977	1980	1985
Hydrocracker throughput required (MB/CD)						
High severity	13.2	19.0	13.4	5.7	6.7	3.1
Medium severity	6.3	0.3	14.7	14.2	13.1	17.5
Subtotal	19.5	19.3	28.1	19.9	19.8	20.6
Less existing throughput (MB/CD)	14.7[a]	19.5[b]	19.5[b]	14.7[a]	19.9[b]	19.9[b]
Subtotal (MB/CD)	4.8	0	8.6	5.2	0	0.7
Less spare capacity (MB/CD)	3.4	0	0	3.4	0	0
New capacity required (MB/CD)	1.4	–	8.6	1.8	–	0.7
Cost of spare capacity[c] ($MM)	5.08	0	0	5.08	0	0
Cost of new capacity[d] ($MM)	1.80	0	11.03	2.31	0	0.90
Total capacity cost ($MM)	6.88	0	11.03	7.39	0	0.90
Cumulative total capacity cost ($MM)	6.88	6.88	17.91	7.39	7.39	8.29

[a] 1973 Calibration throughput, Table E-5

[b] Total capacity available after 1977

[c] Throughput x (1270/0.85)

[d] Throughput x (1090/0.85)

cluster is at high severity (Table E-5). New capacity investment is that for medium severity hydrocracking from Table E-6. In 1977, both scenarios show capacity requirements that exceed the spare capacity available. Thus, the penalty for Scenario C regulations is represented by the difference in cost of new hydrocracker capacity of $0.51 million ($2.31 million for Scenario C less $1.80 million for Scenario A). In 1980, required throughput for both scenarios is slightly less than the total capacity available after 1977 and hence no penalty is incurred. By 1985, Scenario A required $11 million of new hydrocracker capacity versus less than $1 million for Scenario C. Total cumulative onsite cost for Scenario A is $17.91 million while that for Scenario C is $8.29 million, thus showing a net onsite investment credit of $9.62 million for lead regulations. This credit is increased by 40% to account for working capital and offsite costs for a total credit of $13.47 million.

Table E-8 provides an example of the calculation of penalties for flexibility in hydrocracker operating severity for the Texas Gulf cluster, Scenarios A and C. In 1977, Sceanrio A requires 6.3 MB/CD of hydrocracking at medium severity as shown in Table E-7. Since existing capacity is all at high severity (see Table E-5), throughput requirements at medium severity must be charged with the cost of alterations to provide severity flexibility. New capacity construction of 1.4 MB/CD is subtracted from the medium severity throughput required, as new grassroots investment is assumed to be installed at the appropriate severity level. Additional severity flexibility needed is thus 4.9 MB/CD. After 1977 medium severity capacity available is the sum of new capacity construction and severity flexibility provided in 1977.

Similarly, for Scenario C medium severity hydrocracker throughput requirements are those reported in Table E-7, while capacity available is that shown in Table E-5 for 1977 and the capacity installed or upgraded in previous periods for 1980 and 1985. The cost of severity flexibility is the stream day investment from Table E-6, adjusted to a calendar day basis.

Cumulative onsite costs for providing flexibility in hydrocracking severity is $4.23 million in Scenario C compared to $1.38 million in Scenario A. The cumulative onsite severity change penalty for lead regulations in the Texas Gulf cluster is thus $2.85 million, which when

Table E-8. COST OF HYDROCRACKER SEVERITY FLEXIBILITY

Texas Gulf Cluster

	Scenario A			Scenario C		
	1977	1980	1985	1977	1980	1985
Medium severity hydrocracker throughput required (MB/CD)	6.3	0.3	14.7	14.2	13.1	17.5
Less medium severity capacity available (MB/CD)	0[a]	6.3[b]	6.3[b]	–[a]	14.2[b]	14.2[b]
Total	6.3	0	8.4	14.2	0	3.3
Less new capacity construction (MB/CD)	1.4[c]	0	8.6[d]	1.8[c]	0	0.7[d]
Severity flexibility required (MB/CD)	4.9	0	0	12.4	0	2.6
Cost of severity flexibility[e] ($MM)	1.38	0	0	3.50	0	0.73
Cumulative cost of severity flexibility ($MM)	1.38	1.38	1.38	3.50	3.50	4.23

[a] 1973 Existing capacity, Table E-5

[b] Total capacity available after 1977

[c] New capacity built in 1977, Table E-7

[d] New capacity built in 1985, Table E-7

[e] Throughput × (240/0.85)

increased by 40% for working capital and offsite costs, results in a total flexibility penalty of $3.99 million.

The total cumulative penalty or credit, including new capacity investment, the charge for utilization of existing capacity, and costs of providing severity flexibility, is obtained by adding the results of Tables E-7 and E-8. The result is a total hydrocracker investment credit of $9.5 million for lead regulations in the Texas Gulf cluster. The credit to the total U.S. refining industry is found by multiplying this credit by the appropriate scale-up factor in Appendix G.

C. ALKYLATION

Calibration results for utilization of alkylation capacity are compared with existing calendar day capacity in the upper half of Table E-9. All but the Louisiana Gulf and West Coast clusters show calibration requirements exceeding the calculated existing capacity. The penalty for utilization of spare capacity has been taken as the investment cost for new alkylation capacity as shown below:

	Standard Unit, MB/SD	Investment For New Capacity, $/B/SD
Small Midcontinent Cluster	5	2250
All Other Clusters	10	1400

A sample calculation of investment penalties for utilization of existing alkylation capacity is shown for the Louisiana Gulf, Scenarios A and C, in Table E-10; no new capacity is required in this case. Alkylation throughput required in Scenario A is less than existing calibration throughput for all years and hence there is no cost for use of spare capacity. In 1980, Scenario C requires 0.2 MB/CD of spare capacity which is charged at the stream-day investment cost of $1400/B given above and adjusted to a calendar day basis. In 1985, Scenario C, existing throughput represents total capacity available after 1980. Spare capacity available for utilization in 1985 has been reduced by the spare capacity used (and hence charged off) in 1980. Since no new capacity is required in Scenario C, the cumulative spare capacity investment penalty of $.99 million represents total onsite penalties. Including working capital and offsite costs at 40% of onsite investment, the total alkylation-related penalty for lead regulations is $1.39 million for

Table E-9. ALKYLATION AND ISOMERIZATION CAPACITY AVAILABILITY

(MB/CD)

	Cluster model					
	East Coast	Large Midwest	Small Midcont.	Louisiana Gulf	Texas Gulf	West Coast
Alkylation						
1973 Calibration throughput	8.0	12.0	4.9	17.5	17.8	5.5
Existing capacity[a]	7.1	11.4	4.5	20.4	17.7	6.6
Spare capacity available	–	–	–	2.9	–	1.1
Isomerization						
1973 Calibration throughput	–	–	–	–	–	–
Existing capacity-once through[a]	–	–	1.5	–	2.0	–
Spare capacity available	–	–	1.5	–	2.0	–

[a] 85% of Oil and Gas Journal stream day capacity

Table E-10. UTILIZATION OF EXISTING ALKYLATION CAPACITY

Louisiana Gulf Cluster

	Scenario A			Scenario C		
	1977	1980	1985	1977	1980	1985
Alkylation throughput required (MB/CD)	16.9	16.9	16.6	17.4	17.7	18.1
Less existing throughput (MB/CD)	17.5[a]	17.5[a]	17.5[a]	17.5[a]	17.5	17.7[b]
Subtotal (MB/CD)	0	0	0	0	0.2	0.4
Less spare capacity (MB/CD)	–	–	–	–	2.9	2.7[c]
New capacity required (MB/CD)	–	–	–	–	0	0
Cost of spare capacity[d] ($MM)	0	0	0	0	0.33	0.66
Cumulative cost of spare capacity ($MM)	0	0	0	0	0.33	0.99

[a] 1973 Calibration throughput, Table E-9

[b] Total capacity available after 1980

[c] Spare capacity available after 1980

[d] Throughput x (1400/0.85)

the Louisiana Gulf cluster. This figure, multiplied by the appropriate scale-up factor in Appendix G, is the contribution to the total penalty to the U.S. refining industry represented by the Louisiana Gulf cluster.

D. ISOMERIZATION

As shown in the lower half of Table E-9, only two clusters--the Small Midcontinent and the Texas Gulf clusters--had existing isomerization capacity, although calibration results showed no isomerization throughput requirements. Grassroots investment for recycle isomerization is twice the cost of new once through capacity, as shown in Table E-11. All existing capacity is assumed to be once through. The investment for upgrading once through to recycle isomerization and the charge for utilitzation of existing capacity are both equal to initial investment for once through isomerization.

Table E-12 shows the calculation of investment penalties for utilization of isomerization capacity for Scenarios A and C in the Texas Gulf cluster. Scenario A did not require isomerization in any of the three years. In 1977, Scenario C required 2.6 MB/CD of total isomerization capacity, thus utilizing all 2.0 MB/CD of spare capacity in this year. The cost of using spare capacity is the $620/B stream day investment for use of existing once through capacity from Table E-11, multiplied by the 2.0 MB/CD throughput and adjusted to a calendar day basis. The 0.6 MB/CD of new recycle capacity is multiplied by the investment figure in Table E-11 for new recycle capacity, for a cost of $.88 million. In 1980, the existing throughput is that available after 1977 (2.6 MB/CD). Since all spare capacity was utilized and charged in 1977, additional requirements must come from construction of new capacity, which in this case is all recycle isomerization. Scenario C in 1985 requires both new once through and new recycle capacity. New once through capacity required is determined by subtracting the 0.3 MB/CD once through capacity in 1980 from the 7.0 MB/CD once through throughput required in 1985. The cost of new once through isomerization, $4.89 million, is then found by multiplying the required new capacity (6.7 MB/CD) times the stream day investment cost for new once through isomerization given in Table E-11 and adjusted to a calendar day basis. The cost for new recycle isomerization, $3.21 million, is calculated in a similar manner.

Table E-11. ISOMERIZATION INVESTMENT FOR CAPACITY-UTILIZATION AND ONCE THROUGH UPGRADING

	Small Midcontinent cluster		All other clusters	
	Standard unit, MB/SD	Investment $/B/SD	Standard unit, MB/SD	Investment $/B/SD
Existing and new capacity				
– once through	5	1,000	10	620
– recycle	5	2000	10	1,240
Once through upgrading	5	1,000	10	620

Table E-12. COSTS OF ADDITIONAL ISOMERIZATION CAPACITY

Texas Gulf Cluster

	Scenario A			Scenario C		
	1977	1980	1985	1977	1980	1985
Isomerization throughput required (MB/CD)						
– once through	0	0	0	0.3	0.3	7.0
– recycle	0	0	0	2.3	6.1	8.3
Subtotal	0	0	0	2.6	6.4	15.3
Less existing throughput (MB/CD)	0[a]	0[a]	0[a]	0[a]	2.6[b]	6.4[c]
Subtotal (MB/CD)	0	0	0	2.6	3.8	8.9
Less spare capacity (MB/CD)	2.0	2.0	2.0	2.0	–	–
New capacity required (MB/CD)						
Total	0	0	0	0.6	3.8	8.9
– once through	0	0	0	–	–	6.7
– recycle	–	–	–	0.6	3.8	2.2
Cost of spare capacity[d] ($MM)	–	–	–	1.46	–	–
Cost of new capacity ($MM)						
– once through[d]	–	–	–	–	–	4.89
– recycle[e]	–	–	–	0.88	5.54	3.21
Total capacity cost ($MM)	0	0	0	2.34	5.54	8.10
Cumulative total capacity cost ($MM)	0	0	0	2.34	7.88	15.98

[a] 1973 Calibration throughput, Table E-9

[b] Total capacity available after 1977

[c] Total capacity available after 1980

[d] Throughput × (620/0.85)

[e] Throughput × (1240/0.85)

Cumulative onsite cost for use of existing isomerization capacity and construction of new capacity for Scenario C versus A is $15.98 million. Increasing this onsite investment by 40% for offsite costs and working capital gives an isomerization penalty of $22.37 million for lead regulations in the Texas Gulf cluster.

Investment penalties for upgrading once through to recycle isomerization are given in Table E-13 for Scenarios A and C of the Texas Gulf cluster. Scenario A did not use isomerization and thus has a cumulative cost of zero. Recycle isomerization throughput required in Scenario C is taken from Table E-12. Since existing capacity is assumed to be once through isomerization only (Table E-9), there is no recycle capacity available in 1977. Total throughput requirements of 2.3 MB/CD less 0.6 MB/CD new recycle capacity built in 1977 show 1.7 MB/CD of once through isomerization that must be upgraded. This figure is multiplied by the stream-day investment in Table E-11 for once through upgrading and adjusted to a calendar day basis for a cost of $1.24 million. No further once through upgrading is required in 1980 and 1985 so the cumulative onsite investment penalty for Scenario C is $1.24 million. Including working capital and offsite costs at 40% of onsite investment, the penalty for upgrading becomes $1.74 million.

Combining the results of additional capacity utilization from Table E-12 and once through upgrading from Table E-13 gives a total isomerization investment penalty for Scenario C versus A of $24.1 million. Multiplying by the scale up factors of Appendix G gives the contribution to the U.S. industry of the Texas Gulf cluster penalty.

Table E-13. COST OF ONCE THROUGH ISOMERIZATION UPGRADING

Texas Gulf Cluster

	Scenario A			Scenario C		
	1977	1980	1985	1977	1980	1985
Recycle isomerization throughput required (MB/CD)	0	0	0	2.3	6.1	8.3
Less recycle capacity available (MB/CD)	0	0	0	0[a]	2.3[b]	6.1[c]
Total				2.3	3.8	2.2
Less new recycle capacity construction (MB/CD)	0	0	0	0.6[d]	3.8[e]	2.2[f]
Once through upgrading required (MB/CD)	0	0	0	1.7	0	0
Cost of once through upgrading[g] (SMM)	0	0	0	1.24	0	0
Cumulative cost of upgrading (SMM)	0	0	0	1.24	1.24	1.24

[a] 1973 Existing capacity, Table E-9

[b] Total recycle capacity available after 1977

[c] Total recycle capacity available after 1980

[d] New capacity built in 1977, Table E-12

[e] New capacity built in 1980, Table E-12

[f] New capacity built in 1985, Table E-12

[g] Throughput x (620/0.85)

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APPENDIX F

DEVELOPMENT OF CLUSTER MODELS

The U.S. refining industry is composed of nearly 300 individual refineries scattered throughout the country; each is characterized by a unique capacity, processing configuration, and product distribution, often varying significantly from refinery-to-refinery. In developing a refining model for the industry, one could attempt to aggregate all these refineries into a single composite refinery model. Of necessity, such an approach would require averaging the U.S. crude slate, processing sequence and unit size, and product distribution in this single refinery model. Furthermore, since the entire industry is simulated as a single composite refinery, the model would exhibit a higher degree of flexibility in processing configurations and in selective crude/product blending options than the industry could achieve in practice. For these reasons, the single composite refinery model was not used in these studies. On the other hand, it is impractical to attempt a simulation of the U.S. refining industry by the creation of nearly 300 individual computer models, one representing each individual refinery.

Since there are logical regional groupings of major refineries with similar crude supply patterns, processing configurations, and product outputs, a cluster model approach was developed for this study. In this approach, the existing U.S. refining industry was simulated by a relatively small number of cluster refinery models. The results of these individual cluster model studies were composited by using appropriate scale (or weighting) factors to provide a representation of the U.S. refining industry.

To overcome the disadvantages of the single composite model approach, it is necessary that each cluster model crude slate, processing configuration and product outputs closely approximate the specific refineries which the cluster model is intended to represent. As discussed below, this was achieved by selecting six geographic regions in the United States having refineries with similar characteristics, and creating one cluster model for each region. Hence, a total of six cluster models was used as an appropriate balance between one single composite refinery model and nearly 300 individual refinery models.

Furthermore, it is important that each of these cluster refinery models represent, as closely as possible, a realistic mode of operation of the actual refineries being simulated. Specifically, processing units should be of normal commercial size and the plants should be allowed normal flexibility in regard to raw material selection and product mix.

Also, to allow verification of these objectives, each cluster model should be calibrated against historical operating data of the refinery being simulated. Since operating data is confidential for any given refinery, it was agreed that input/output data, energy consumptions, and plant operating data be supplied from government and industry as an aggregate of three specific refineries selected to comprise each cluster model. By using three, it would be impossible to determine competitive proprietary data for any single refinery.

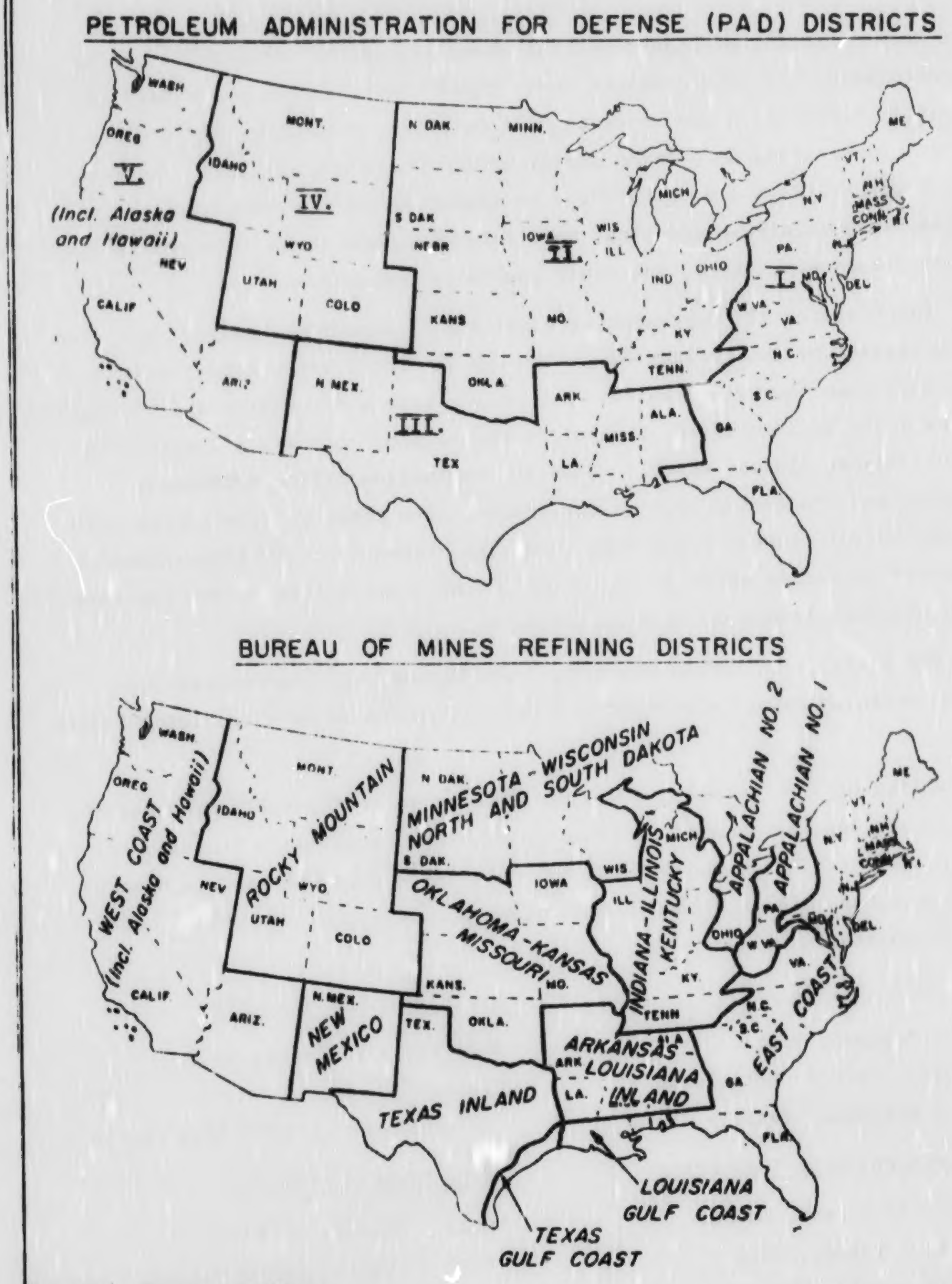
The initial task in this program was the identification of the regions used to represent the U.S. industry and the selection of the specific three individual refineries which would make up each cluster model. An ad hoc industry task force comprised of API and NPRA representatives played a major role in this effort.

A. SELECTION OF CLUSTER MODELS

The selection of the geographic regions to be simulated in the cluster models required definition of several guidelines, summarized below.

PAD Districts I and V (Figure F-1) have sufficient crude capacity of common characteristics that they can be simulated by one cluster model for each district.

PAD District III, which represents about 40% of the U.S. total refining capacity, was simulated by two models because of its overall importance and because two types of refining configurations were identified. The Louisiana


PETROLEUM ADMINISTRATION FOR DEFENSE (PAD) DISTRICTS

BUREAU OF MINES REFINING DISTRICTS

V. (Incl. Alaska and Hawaii)
IV.
III.
II.
I.
WASH
OREG
IDAHO
MONT.
N DAK
MINN.
S DAK
NEBR
WYO
NEV
UTAH
COLO
CALIF
KANS
MO.
IOWA
WIS
ILL
MICH
IND
OHIO
KY
TENN.
ARIZ
N MEX
OKLA
TEX
ARK
LA.
MISS.
ALA.
GA
FLA
S.C.
N.C.
VA.
W VA
PA.
MD
DEL
NJ
NY
VT
NH
MASS
CONN
R.I.
ME
WEST COAST (Incl. Alaska and Hawaii)
ROCKY MOUNTAIN
MINNESOTA - WISCONSIN NORTH AND SOUTH DAKOTA
OKLAHOMA - KANSAS MISSOURI
INDIANA - ILLINOIS KENTUCKY
APPALACHIAN NO. 2
APPALACHIAN NO. 1
NEW MEXICO
TEXAS INLAND
ARKANSAS - LOUISIANA INLAND
EAST COAST
LOUISIANA GULF COAST
TEXAS GULF COAST

Source: Bureau of Mines.

Figure F-1. GEOGRAPHIC REGIONS CONSIDERED IN DEVELOPMENT OF CLUSTER MODELS.

Gulf Coast district of Figure F-1 (or about 1/3 of the total Gulf) can be characterized by a single source sweet crude slate, a high percentage of catalytic cracking, a low percentage of catalytic reforming, and product outputs emphasizing both major energy products and specialties. The Texas Gulf Coast district can be typified by higher sulfur and more varied crude slates, less catalytic cracking, more reforming, and heavy involvement in petrochemicals, lubes, and other specialty operations.

PAD District II, representing about 28% of domestic capacity, was also characterized by two refineries. Although its total crude capacity is somewhat less than the Gulf, two separate models with quite distinct characteristics could be identified to simulate the region. One was a large (100+ MB/CD) Midwest cluster model simulating the Indiana/Illinois/Kentucky district and processing high sulfur crudes. The other was a moderate size (50-100 MB/CD) cluster model simulating the Midcontinent (Oklahoma/Kansas/Missouri) operation which is characterized by lower sulfur crudes and very low production levels of residual products.

PAD District IV (Rocky Mountains, see Figure F-1) represented too little refining capacity (<5% of total) to be included as a specific cluster model.

A list of U.S. refineries was prepared representing refineries suitable for the aggregation program outlined above. The final identification of the specific refineries comprising each cluster was jointly agreed upon by the contractor, EPA, and the ad hoc industry task force. The final selection is indicated below:

Texas Gulf (PAD III)

Exxon - Baytown, Texas
Gulf Oil - Port Arthur, Texas
Mobil - Beaumont, Texas

Small Midcontinent (PAD II)

Skelly - El Dorado, Kansas
Gulf Oil - Toledo, Ohio
Champlin - Enid, Oklahoma

Louisiana Gulf (PAD III)

Gulf Oil - Alliance, LA
Shell Oil - Norco, LA
Cities Service - Lake Charles, LA

Large Midwest (PAD II)

Mobil - Joliet, Illinois
Union - Lemont, Illinois
Arco - East Chicago, Illinois

East Coast (PAD I)	West Coast (PAD V)
Arco - Philadelphia, PA	Mobil - Torrance, California
Sun Oil - Marcus Hook, PA	Arco - Carson, California
Exxon - Linden, New Jersey	Socal - El Segundo, California

B. COMPARISON OF CLUSTER MODEL TO PAD DISTRICT

The major factor in the original selection of the three refineries comprising each cluster was the processing configuration. Tables F-1 to F-6 provide detailed processing information for the three selected refineries in each cluster for January 1, 1973, and January 1, 1974, as presented in the Oil and Gas Journal annual refining surveys. Table F-7 compares the key processing configuration for each cluster refinery to the corresponding PAD total. In PAD III the Texas and Louisiana Gulf refineries bracket the PAD average for coking and hydrocracking. They are slightly low on reforming and high on catalytic cracking and alkylation. Since the clusters are intended to represent large refiners that produce high yields of gasoline rather than small specialty plants maximizing asphalt and/or lubes, this is to be expected. In PAD II, the coking and catalytic reforming capacity for the cluster models bracket the PAD average. Catalytic cracking is high for this PAD region; however, since the cluster refineries contain no hydrocracking, the composite of cracking conversion operations checks well against the PAD average.

In PAD I, the East Coast cluster contained no coking, although it did possess a higher percentage of catalytic cracking and hydrocracking than the PAD average. It was considered useful to have at least one cluster model that did not contain coking, as this is characteristic of many U.S. plants. The West Coast cluster refineries also exhibited slightly greater process unit intakes than the PAD average with the exception of alkylation.

Once the specific refineries comprising each cluster were identified, cluster input/output data for the year 1973 was requested from the Bureau of Mines (BOM). This information was then tabulated and compared to the average of the BOM data for the entire PAD district to determine if the specific crude slates and product output patterns for the cluster refineries were representative of the PAD average.

Table F-1. TEXAS GULF CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity,[a] 1974				Unit capacity,[a] 1973				Unit capacity,[a,b]	
	Exxon Baytown, Texas	Gulf Port Arthur, Texas	Mobil Beaumont, Texas	1974 Average	Exxon Baytown, Texas	Gulf Port Arthur, Texas	Mobil Beaumont, Texas	1973 Average	1973/1974 Average	% Crude
Crude capacity, B/CD	430,000	312,100	325,000	346,700	350,000	312,100	335,000	332,367	339,034	
	420,000	319,000	335,000	358,000	365,000	319,000	350,000	344,667	351,333	
Vacuum dist.	180,000	147,400	103,000	143,467	150,000	147,400	103,000	133,467	138,467	39.4
Thermal										
–Visb.							12,000	4,000	2,000	0.6
–Fluid coke										
–Delayed coke		30,000	33,000	21,000		30,000	33,000	21,000	21,000	6.0
Other										
Catalytic cracking	124,000	120,000	95,000	113,000	135,000	120,300	95,000	116,667	114,834	32.7
Catalytic reforming	88,000	65,000	94,000	82,333	88,000	65,000	94,000	82,333	82,333	23.4
Hydrocracking										
–Dist.	20,000	15,000	29,000	21,333	20,000	15,000	29,000	21,333	21,333	6.1
–Residual										
–Lubes										
–Other										
Hydrofining										
–Hvy gas oil	48,000			16,000	53,000			17,667	16,834	4.8
–Resid. visb.										
–Cat feed & cycle										
–Distillate		65,000		21,667		65,000		21,667	21,667	6.2
–Other										
Hydrotreat										
–Reform feed	90,000	65,000	83,000	79,333	90,000	65,000	83,000	79,333	79,333	22.6
–Naphtha	15,000			5,000	32,000			10,667	7,833	2.2
–Olef/Arom sat		1,200		400		1,200		400	400	0.1
-S.R. Distill.										
–Lubes	41,000	13,900		18,300	39,500	13,900		17,800	18,050	5.1
Other dist.	109,000		42,000	50,333	84,000		42,000	42,000	46,167	13.1
Other	8,500[c]			2,833	8,500			2,833	2,833	0.8
Alkylation	26,000	20,000	16,500	20,833	26,000	20,000	16,500	20,833	20,833	5.9
Arom/Isom										
–BTX		2,700		900		2,700		900	900	0.3
–HDA										
-Cyclohex		2,500		833		2,500		833	833	0.2
–C4 Feed										
–C5 Feed		7,200		2,400		7,200		2,400	2,400	0.7
–C5/C6 Feed										
Lubes	25,000	13,200	8,800	15,667	25,000	13,200	8,800	15,667	15,667	4.4
Asphalt	12,000		100	4,033	12,000		100	4,033	4,033	1.1
Coke–tons/day		1,390	1,200	863		1,390	1,200	863	863	–

[a] B/SD unless otherwise noted.

[b] Used in cluster model.

[c] Solvents.

Reference: *Oil and Gas Journal,* April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table F-2. LOUISIANA GULF CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity,[a] 1974				Unit capacity,[a] 1973				Unit capacity,[a,b]	
	Gulf Alliance, La.	Shell Norco, La.	Citgo L. Chas., La.	1974 Average	Gulf Alliance, La.	Shell Norco, La.	Citgo L. Chas., La.	1973 Average	1973/ 1974 Average	% Crude
Crude capacity, B/CD	180,400	240,000	268,000	229,467	174,000	240,000	240,000	218,000	223,734	
	186,000	250,000	N.R.	–	180,000	250,000	245,000	225,000	–	
Vacuum dist.	55,000	90,000	60,000	68,333	54,000	90,000	78,000	74,000	71,167	30.8
Thermal										
–Visb.										
–Fluid coke										
–Delayed coke	16,000	18,000	28,000	20,667	16,000	17,000	25,000	19,333	20,000	8.7
–Other										
Catalytic cracking	78,000	95,000	125,000	99,333	75,000	85,000	112,500	90,833	95,000	41.1
Catalytic reforming	37,500	41,500	46,000	41,667	37,500	43,000	39,000	39,833	40,750	17.6
Hydrocracking										
–Dist.		28,000		9,333		29,400		9,800	9,566	4.1
–Residual							6,000	2,000	1,000	0.4
–Lubes										
–Other										
Hydrofining										
–Residual			6,000	2,000					1,000	0.4
–Hvy gas oil										
–Resid. visb.										
–Cat feed & cycle	16,000	25,000	30,000	23,667	16,000	25,000		13,667	13,667	8.1
–Distillate	22,000			7,333	22,000			7,333	7,333	3.2
–Other										
Hydrotreat										
–Reform feed	41,000		46,000	29,000	41,000		16,300	19,100	24,050	10.4
–Naphtha		26,000		8,667		26,000	11,200	12,400	10,534	4.6
–Olef/Arom sat										
–S.R. Distill.			14,000	4,667					2,333	1.0
–Lubes										
–Other dist.										
–Other							24,000	8,000	4,000	1.7
Alkylation	28,400	14,100	35,300	25,933	28,400	14,100	26,000	22,833	24,383	10.6
Arom/Isom										
–BTX	11,100			3,700	11,100			3,700	3,700	1.6
–HDA	5,400			1,800	5,400			1,800	1,800	0.8
–Cyclohex										
–C4 Feed										
–C5 Feed										
C5/C6 Feed										
Lubes			7,000	2,333			10,000	3,333	2,833	1.2
Asphalt		6,000		2,000		6,000		2,000	2,000	0.9
Coke–tons/day	840	900	1,000	913	840	900	895	878	896	–

[a] B/SD unless otherwise noted.

[b] Used in cluster model.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table F-3. LARGE MIDWEST CLUSTER PROCESS CONFIGURATION

Unit type	Unit capacity,[a] 1974				Unit capacity,[a] 1973				Unit capacity,[a,b]	
	Mobil Joliet, Illinois	Union Lemont, Illinois	Arco E. Chic., Illinois	1974 Average	Mobil Joliet, Illinois	Union Lemont, Illinois	Arco E. Chic., Illinois	1973 Average	1973/ 1974 Average	% Crude
Crude capacity, B/CD	175,000	152,000	126,000	151,000	160,000	140,000	135,000	145,000	148,000	
Vacuum dist.	186,000	N.R.	140,000		164,000	N.R.	140,000			
Thermal	82,000	55,000	70,000	69,000	72,500	55,000	70,000	65,833	67,417	43.4
–Gas oil										
–Visb						19,000		6,333	3,167	2.0
–Fluid coke										
–Delayed coke	28,000	19,500		15,833	28,000			9,333	12,583	10.1
–Other										
Catalytic cracking	66,000	52,000	48,000	55,333	66,000	50,000	48,000	54,667	55,000	35.4
Catalytic reforming	46,200	32,000	20,000	32,733	46,200	32,000	20,000	32,733	32,733	21.1
Hydrofining										
–Hvy gas oil										
–Resid. visb.										
–Cat feed & cycle										
–Distillate	69,000		25,000	31,333			25,000	8,333	19,833	12.7
–Other										
Hydrotreat										
–Reform feed	67,000	32,000	20,000	39,667	53,000	32,000	20,000	35,000	37,334	24.0
–Naphtha		2,700	2,000	1,567		2,000	2,000	1,333	1,450	.9
–Olef/Arom sat		4,500		1,500		5,300		1,767	1,634	1.1
–S.R. Distill.		7,000		2,333		7,000		2,333	2,333	1.5
–Lubes										
–Other dist.		34,500		11,500	54,000			18,000	14,750	12.9
–Other		2,500[c]		833		37,000		12,333	6,583	4.2
Alkylation	22,000	12,800	6,000	13,600	18,000	16,000	6,000	13,333	13,467	8.6
Arom/Isom										
–BTX		3,300		1,100		3,200		1,067	1,084	.7
–HDA										
–Cyclohex										
–C4 Feed										
–C5 Feed										
–C5/C6 Feed										
Lubes										
Asphalt		2,000	10,400	4,133			10,400	3,467	3,800	2.4
Coke–tons/day	1,700	1,000		900	1,700	1,000		900	900	–

[a] B/SD unless otherwise noted.

[b] Used for cluster model.

[c] Benzene concentrate.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table F-4. SMALL MIDCONTINENT CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity,[a] 1974				Unit capacity,[a] 1973				Unit capacity,[a,b]	
	Skelly El Dorado, Kan.	Gulf Toledo, Ohio	Champlin Enid, Okla.	1974 Average	Skelly El Dorado, Kan.	Gulf Toledo, Ohio	Champlin Enid, Okla.	1973 Average	1973/ 1974 Average	% Crude
Crude capacity,B/CD	73,700	50,300	49,500	57,833	67,000	48,800	46,000	54,600	56,217	
	75,000	51,000	52,000	59,333	70,000	50,000	50,000	56,667	58,000	
Vacuum dist	23,000	12,500	18,000	17,833	23,000	12,300	24,000	19,767	18,800	32.4
Thermal										
–Visb.										
–Fluid coke										
–Delayed coke	9,800		3,700	4,500	9,800		4,000	4,600	4,550	7.8
–Other										
Catalytic cracking	30,000	20,000	19,500	23,167	30,000	18,500	19,000	22,500	22,834	39.4
Catalytic reforming	21,500	11,000	15,000	15,833	20,000	10,500	15,000	15,167	15,500	26.7
Hydrofining										
–Hvy gas oil										
–Resid. visb.										
–Cat feed & cycle										
–Distillate		5,000		1,667		5,000		1,667	1,667	2.9
–Other										
Hydrotreat										
–Reform feed	23,000	11,000	20,400	18,133	23,000	10,500	15,000	16,167	17,150	29.6
–Naphtha										
–Olef/Arom sat	4,300			1,433	4,300			1,433	1,433	2.5
–S.R. Distill.										
–Lubes										
–Other dist.										
–Other							5,000[c]	1,667	834	1.4
Alkylation	6,000	5,500	4,500	5,333	6,000	5,100	4,400	5,167	5,250	9.1
Arom/Isom										
–BTX	1,400			467	1,400			467	467	0.8
–HDA										
–Cyclohex										
–C4 Feed										
–C5 Feed										
–C5/C6 Feed			6,000	2,000			5,000	1,667	1,834	3.2
Lubes			1,100	367			1,200	400	384	0.7
Asphalt		2,000	1,400	1,133	3,000	2,000	2,000	2,333	1,733	3.0
Coke–tons/day	500		165	222	500		158	219	221	–

[a] B/SD Unless otherwise noted.

[b] Used for cluster model.

[c] Isom feed.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table F-5. EAST COAST CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity,[a] 1974				Unit capacity,[a] 1973				Unit capacity,[a,b]	
	Arco Phil., Pa.	Sun Marcus Hook, Pa.	Exxon Linden, N.J.	1974 Average	Arco Phil., Pa.	Sun Marcus Hook, Pa.	Exxon Linden, N.J.	1973 Average	1973/1974 Average	% Crude
Crude capacity, B/CD	185,000	165,000	275,000	208,333	160,000	163,000	255,000	192,667	200,500	
	195,000	180,000	286,000	220,333	165,000	180,000	268,000	204,333	212,333	
Vacuum dist.	57,000	48,000	143,000	82,667	83,000	48,000	140,000	90,333	86,500	40.7
Thermal										
–Visb.										
–Fluid coke										
–Delayed coke										
–Other										
Catalytic cracking		75,000	120,000	65,000	36,000	75,000	125,000	78,667	71,834	33.8
Catalytic reforming	60,000	45,000	42,000	49,000	60,000	43,000	46,000	49,667	49,334	23.2
Hydrocracking										
–Dist.	30,000			10,000	30,000			10,000	10,000	4.7
–Residual										
–Lubes										
–Other										
Hydrofining										
–Hvy gas oil	32,000		50,000	27,333			50,000	16,667	22,000	10.4
–Resid, visb.										
–Cat feed & cycle										
–Distillate	41,000			13,667	34,000			11,333	12,500	5.9
–Other										
Hydrotreat										
–Reform feed	54,000	35,000	42,000	43,667	53,000	35,000	46,000	44,667	44,167	20.8
–Naphtha			14,000	4,667			14,000	4,667	4,667	2.2
–Olef/Arom sat										
–S.R. Distill										
–Lubes		10,000		3,333		10,000		3,333	3,333	1.6
–Other dist.			39,000	13,000			37,000	12,333	12,667	6.0
–Other		10,000[c]		3,333		16,000[c]		5,333	4,333	2.0
Alkylation		12,000	8,500	6,833	7,000	12,000	10,700	9,900	8,367	3.9
Arom/Isom										
BTX		5,300		1,767		5,300		1,767	1,767	.8
–HDA										
–Cyclohex										
C4 Feed										
C5 Feed										
–C5/C6 Feed										
Lubes		17,000		5,667		17,000		5,667	5,667	2.7
Asphalt	19,500	12,000	46,000	25,833	17,000	12,000	46,000	25,000	25,416	12.0
Coke—tons/day										

[a] B/SD unless otherwise noted.

[b] Used for cluster model.

[c] Furnace oil.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table F-6. WEST COAST CLUSTER PROCESSING CONFIGURATION

	Unit capacity,[a] 1974				Unit capacity,[a] 1973				Unit capacity,[a,b]	
Unit type	Mobil Torrance, Calif.	Arco Carson, Calif.	Socal El Segundo, Calif.	1974 Average	Mobil Torrance, Calif	Arco Carson, Calif	Socal El Segundo, Calif.	1973 Average	1973/ 1974 Average	% Crude
Crude capacity	123,500 130,000	165,000 173,000	230,000 N.R.	172,833	123,500 130,000	165,000 173,000	N.R. 220,000	174,333		
Vacuum dist.	95,000	93,000	103,000	97,000	95,000	93,000	103,000	97,000	97,000	54.5
Thermal										
–Gas oil		12,500		4,167		23,000		7,667	5,917	3.3
–Visb.	16,000	42,000		19,333	16,000	37,000		17,667	18,500	10.4
–Fluid coke										
–Delayed coke	46,640	30,000	54,000	43,547	46,640	25,500	50,000	40,713	42,130	23.7
–Other										
Catalytic cracking	56,000	57,000	43,500	52,167	56,000	57,000	40,000	51,000	51,584	29.0
Catalytic reforming	36,000	34,000	60,000	43,333	36,000	32,000	62,000	43,333	43,333	24.3
Hydrocracking										
–Dist.	18,000	19,700	49,000	28,900	18,000	17,000	45,000	26,667	27,784	15.6
–Residual										
–Lubes										
–Other										
Hydrofining										
–Hvy gas oil										
–Resid. visb.										
–Cat feed & cycle										
–Distillate		18,000		6,000		18,000		6,000	6,000	3.4
–Other										
Hydrotreat										
–Reform feed	23,000	34,000	40,000	32,333	23,000	32,000	40,000	31,667	32,000	18.0
–Naphtha		18,000		6,000					3,000	1.7
–Olef/Arom sat	16,000			5,000	16,000	18,000		11,000	8,000	4.5
–S.R. Distill.			12,000	4,000			12,000	4,000	4,000	2.2
–Lubes										
–Other dist.	25,000			8,333	23,000			7,667	8,000	4.5
–Other			18,000[c]	6,000			18,000	6,000	6,000	3.4
Alkylation	10,500	7,200	5,900	7,867	10,500	7,200	5,400	7,700	7,784	4.4
Arom/Isom										
–BTX		2,490		830		2,490	1,500	1,330	1,080	0.7
–HDA										
–Cyclohex										
–C4 Feed			1,500	500					250	0.1
–C5 Feed										
–C5/C6 Feed										
Lubes										
Asphalt			8,300	2,767			8,300	2,767	2,767	1.6
Coke–tons/day	2,800	1,800	2,200	2,267	2,800	1,650	2,200	2,217	2,242	–

[a] B/SD unless otherwise noted.

[b] Used for cluster model.

[c] Jet fuel.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table F-7. SUMMARY OF MAJOR REFINERY PROCESSING UNITS

(percentage of crude capacity)

Processing unit	Texas Gulf cluster	La. Gulf cluster	PAD III average	Large Midwest cluster	Small Mid-continent cluster	PAD II average	East Coast cluster	PAD I average	West Coast cluster	PAD V average
Catalytic reforming	23	18	24	21	27	22	23	21	24	23
Catalytic cracking	33	41	31	35	39	34	34	33	29	24
Hydrocracking	6	4	5	0	0	4	5	3	16	15
Alkylation	6	11	6	9	9	7	4	4	4	5
Delayed coking	6	9	8	10	8	9	0	7	24	20

Tables F-8 through F-11 provide the comparison of product outputs. In general, the cluster refineries exhibited higher yields of gasoline and distillate fuel oil (except for East Coast) and lower yields of residual fuel oil (except for West Coast) than the corresponding PAD averages. However, in no cases were the deviations deemed to be of sufficient magnitude to change the cluster make-up.

Tables F-12 through F-15 provide the comparison of the crude slate for each cluster with the crude slate for the PAD district. PAD Districts II and III showed excellent agreement. There were variations in PAD I, however the varying crudes were of equivalent quality. For example, the subtotal of African crudes (light, low sulfur) checked very well and the combined subtotals of Middle East and South American crudes (heavy, high sulfur) also showed excellent agreement. Thus, the major discrepency was the replacement of mixed Canadian crude with domestic supplies (primarily from Texas) which should not appreciably change average crude quality.

In PAD V, Canadian crude is replaced by approximately equal quantities of Middle East and Far East crudes (low sulfur) in the cluster model which, again does not reflect a major change in crude quality.

Table F-8. COMPARISON OF PRODUCT OUTPUT OF EAST COAST CLUSTER TO P.A.D. DISTRICT I, 1973

	Total P.A.D. I		East Coast cluster	
Outputs	MBPY	% of Total	MBPY	% of Total
Gasoline total	272,932	46.58	114,904	50.0
Jet fuel				
Naphtha-type	2,726	.47	1,730	.75
Kerosene-type	11,918	2.03	5,956	2.59
Ethane	58	.01	58	.03
LPG				
For fuel use	14,364	2.45	7,369	3.21
For chemical use	6,394	1.09	4,524	1.97
Kerosene	7,009	1.20	3,711	1.61
Distillate fuel	147,003	25.09	49,818	21.68
Residual fuel	52,258	8.92	14,053	6.12
Petrochemical feedstocks				
Still gas	942	.16	883	.38
Naphtha – 400°	4,932	.84	1,485	.65
Other	768	.13	29	.01
Special naphthas	391	.07	13	.01
Lube oil, total	12,081	2.06	5,074	2.21
Wax	1,433	.24	333	.14
Coke (marketable)	13,627	2.33	0	0
Asphalt	36,416	6.21	19,856	8.64
Road oil	706	.12	0	0
Total	585,958	100.00	229,796	100.0

Table F-9. COMPARISON OF PRODUCT OUTPUT OF MIDCONTINENT CLUSTERS TO P.A.D. DISTRICT II, 1973

Outputs	P.A.D. II		Small Midcontinent cluster		Large Midwest cluster	
	MBPY	% of Total	MBPY	% of Total	MBPY	% of Total
Gasoline total	728,246	54.32	40,991	59.95	89,467	55.42
Jet fuel						
Naphtha-type	11,937	.89	472	.69	143	.09
Kerosene-type	50,788	3.79	911	1.33	2,297	1.42
Ethane	520	.04	520	.76	0	0
LPG						
For fuel use	25,323	1.89	1,586	2.32	3,489	2.16
For chemical use	4,248	.32	653	.96	0	0
Kerosene	19,887	1.48	38	.06	1,813	1.12
Distillate fuel	297,796	22.21	17,374	25.41	44,678	27.67
Residual fuel	71,120	5.31	252	.37	8,094	5.01
Petrochemical feedstocks						
Still gas	2,671	.20	70	.10	0	0
Naphtha – 400°	6,572	.49	1,535	2.25	1,025	.63
Other	2,857	.21	185	.27	0	0
Special naphthas	6,106	.46	7	.01	2,248	1.39
Lube oil, total	10,725	.80	370	.54	0	0
Wax	1,194	.09	0	0	0	0
Coke (marketable)	38,873	2.90	1,430	2.09	4,024	2.49
Asphalt	57,637	4.30	1,976	2.89	2,013	1.25
Road oil	4,104	.31	0	0	2,154	1.33
Total	1,340,605	100.01	68,370	100.0	161,445	99.98

Table F-10. COMPARISON OF PRODUCT OUTPUT OF GULF COAST CLUSTERS TO P.A.D. DISTRICT III, 1973

Outputs	Total P.A.D. III MBPY	Total P.A.D. III % of Total	Louisiana Gulf cluster MBPY	Louisiana Gulf cluster % of Total	Texas Gulf cluster MBPY	Texas Gulf cluster % of Total
Gasoline total	979,079	48.53	130,086	51.08	181,351	48.25
Jet fuel						
Naphtha-type	27,693	1.37	807	.32	3,009	.80
Kerosene-type	114,173	5.66	20,295	7.97	25,115	6.68
Ethane	8,108	.40	687	.27	1,266	.34
LPG						
For fuel use	35,507	1.76	8,644	3.39	5,428	1.44
For chemical use	23,219	1.15	2,623	1.03	4,350	1.16
Kerosene	49,003	2.43	6,021	2.36	8,429	2.24
Distillate fuel	439,979	21.81	69,914	27.45	88,491	23.54
Residual fuel	88,455	4.38	5,856	2.30	17,170	4.57
Petrochemical feedstocks						
Still gas	7,773	.39	0	0	776	.21
Naphtha – 400°	40,298	2.00	42	.02	6,623	1.76
Other	56,170	2.78	3,464	1.36	2,684	.71
Special naphthas	21,010	1.04	0	0	7,231	1.92
Lube oil, total	40,099	1.99	0	0	17,502	4.66
Wax	3,082	.15	0	0	550	.15
Coke (marketable)	42,436	2.10	4,512	1.77	4,380	1.17
Asphalt	41,433	2.05	1,701	.67	1,536	.41
Road oil	64	.003	0	0	0	0
Total	2,017,581	99.993	254,652	99.99	375,891	100.01

Table F-11. COMPARISON OF PRODUCT OUTPUT OF WEST COAST CLUSTER TO P.A.D. DISTRICT V, 1973

Outputs	Total P.A.D. V		West Coast cluster	
	MBPY	% of Total	MBPY	% of Total
Gasoline total	335,285	44.50	74,667	40.96
Jet fuel				
Naphtha-type	20,148	2.67	3,594	1.97
Kerosene-type	66,202	8.79	21,059	11.55
Ethane	508	.07	508	.28
LPG				
For fuel use	12,202	1.62	4,386	2.41
For chemical use	4,139	.55	1,523	.84
Kerosene	1,319	.18	184	.10
Distillate fuel	102,599	13.62	23,891	13.11
Residual fuel	132,900	17.64	33,457	18.35
Petrochemical feedstocks				
Still gas	881	.12	0	0
Naphtha – 400°	5,362	.71	4,271	2.34
Other	3,132	.42	358	.20
Special naphthas	5,241	.70	1,300	.71
Lube oil, total	5,450	.72	381	.21
Wax	961	.13	0	0
Coke (marketable)	33,371	4.43	10,486	5.75
Asphalt	22,013	2.92	2,199	1.21
Road oil	1,682	.22	16	.01
Total	753,385	100.01	182,280	100.00

Table F-12. COMPARISON OF CRUDE INPUT OF EAST COAST CLUSTER TO P.A.D. DISTRICT I, 1973

Crude receipts from:	Total P.A.D. I		East Coast cluster	
	MBPY	% of Total	MBPY	% of Total
P.A.D. I	16,100	2.92	2,594	1.31
P.A.D. II	12,277	2.23	–	–
P.A.D. III				
Louisiana	14,002	2.54	3,346	1.69
Texas and others	43,318	7.85	38,075	19.19
P.A.D. IV	–	–	–	–
Total domestic	85,697	15.53	44,015	22.18
Africa				
Algeria	37,289	6.76	25,248	12.72
Angola	13,884	2.52	–	–
Egypt	5,074	.92	–	–
Libya	26,350	4.78	16,121	8.12
Nigeria	119,281	21.62	29,430	14.83
Tunisia	4,030	.73	3,733	1.88
Subtotal Africa	205,908	37.32	74,532	37.56
Middle East				
Iran	47,496	8.61	3,905	1.97
Iraq	343	.06	–	–
Israel	–	–	–	–
Kuwait	12,665	2.30	–	–
Qatar	–	–	–	–
Saudi Arabia	38,153	6.91	6,036	3.04
United Arab Emirates	12,977	2.35	5,676	2.86
Subtotal Middle East	111,634	20.23	15,617	7.87
South America				
Bolivia	–	–	–	–
Colombia	484	.09	–	–
Ecuador	377	.07	–	–
Mexico	–	–	–	–
Trinidad	4,454	.81	1,772	.89
Venezuela	96,736	17.53	61,344	30.91
Subtotal S. America	102,051	18.50	63,116	31.81
Far East				
Indonesia	2,532	.46	1,165	.59
Malaysia	–	–	–	–
Subtotal Far East	2,532	.46	1,165	.59
Canada	43,949	7.97	–	–
Total foreign	466,074	84.47	154,430	77.82
Total crude	551,771	100.00	198,445	100.00

Table F-13. COMPARISON OF CRUDE INPUT TO MID CONTINENT CLUSTER TO P.A.D. DISTRICT II, 1973

	Total P.A.D. II		Small Midcontinent cluster		Large Midwest cluster	
Crude receipts from:	MBPY	% of Total	MBPY	% of Total	MBPY	% of Total
P.A.D. I						
P.A.D. II	337,673	26.52	40,985	67.72	21,405	12.80
P.A.D. III						
Louisiana	183,950	14.44	2,259	3.73	19,354	11.57
Texas and others	391,308	30.73	5,056	8.35	81,478	48.72
P.A.D. IV	100,160	7.87	1,481	2.45	14,457	8.64
Total domestic	1,013,091	79.55	49,781	82.25	136,694	81.73
Africa						
Algeria	1,438	.11	–	–	–	–
Angola	–	–	–	–	–	–
Egypt	222	.02	–	–	–	–
Libya	5,546	.44	–	–	–	–
Nigeria	4 618	.36	–	–	238	.14
Tunisia	–	–	–	–	–	–
Subtotal Africa	11,824	.93	–	–	238	.14
Middle East						
Iran	6,709	.53	–	–	–	–
Iraq	–	–	–	–	–	–
Israel	–	–	–	–	–	–
Kuwait	–	–	–	–	–	–
Qatar	653	.05	–	–	–	–
Saudi Arabia	17,509	1.37	–		4,291	2.57
United Arab Emirates	639	.05	–	–	–	–
Subtotal Middle East	25,510	2.00	–	–	4,291	2.57
South America						
Bolivia	596	.05	–	–	–	–
Colombia	–	–	–	–	–	–
Ecuador	238	.02	–	–	–	–
Mexico	–	–	–	–		–
Trinidad	4,077	.32	–	–	–	–
Venezuela	1,050	.08	–	–	–	–
Subtotal S. America	5,961	.47	–	–	–	–
Far East						
Indonesia	–	–	–	–	–	–
Malaysia	–	–	–	–	–	–
Subtotal Far East	–	–	–	–	–	–
Canada	217,073	17.05	10,744	17.75	26,022	15.56
Total foreign	260,368	20.45	10,744	17.75	30,551	18.27
Total crude	1,273,459	100.00	60,525	100.00	167,245	100.00

Table F-14. COMPARISON OF CRUDE INPUT OF GULF COAST CLUSTERS TO P.A.D.
District III, 1973

Crude Receipts From:	Total P.A.D. III		Louisiana Gulf cluster		Texas Gulf cluster	
	MBPY	% of Total	MBPY	% of Total	MBPY	% of Total
P.A.D. I	–	–	–	–	–	–
P.A.D. II	–	–	2,395	1.03	–	–
P.A.D. III						
Louisiana	629,470	34.73	185,654	80.13	18,306	5.37
Texas and others	1,037,412	57.24	40,552	17.50	281,252	82.52
P.A.D. IV	–	–	–	–	–	–
P.A.D. V						
California	–	–	–	–	–	–
Other states	–	–	–	–	–	–
Total domestic	1,666,882	91.96	228,601	98.66	299,558	87.89
Africa						
Algeria	4,892	.27	–	–	–	–
Angola	3,869	.21	–	–	3,869	1.14
Egypt	–	–	–	–	–	–
Libya	16,689	.92	214	.09	50	.01
Nigeria	39,788	2.20	827	.36	10,213	3.00
Tunisia	2,511	.14	–	–	–	–
Subtotal Africa	67,479	3.74	1,041	.45	14,132	4.15
Middle East						
Iran	11,041	.61	–	–	3,666	1.08
Iraq	671	.04	–	–	–	–
Israel	309	.02	–	–	–	–
Kuwait	340	.02	–	–	–	–
Qatar	910	.05	–	–	–	–
Saudi Arabia	28,345	1.56	910	.39	15,732	4.62
United Arab Emirates	958	.05	546	.24	–	–
Subtotal Middle East	42,574	2.35	1,456	.63	19,398	5.70
South America						
Bolivia	295	.02	–	–	–	–
Colombia	–	–	–	–	–	–
Ecuador	566	.03	–	–	–	–
Mexico	489	.03	–	–	489	.14
Trinidad	13,208	.73	189	.08	–	–
Venezuela	19,108	1.05	263	.11	7,257	2.13
Subtotal S. America	33,666	1.86	452	.20	7,746	2.27
Far East						
Indonesia	1,665	.09	161	.07	–	–
Malaysia	–	–	–	–	–	–
Subtotal Far East	1,665	.09	161	.07	–	–
Canada	–	–	–	–	–	–
Total foreign	145,654	8.04	3,110	1.33	41,276	12.11
Total crude	1,812,536	100.00	229,316	99.99	340,834	100.00

Table F-16. COMPARISON OF CRUDE INPUT TO WEST COAST CLUSTER
P.A.D. DISTRICT V, 1973

	Total P.A.D. V		West Coast cluster	
Crude receipts from:	**MBPY**	**% of Total**	**MBPY**	**% of Total**
P.A.D. I	–	–	–	–
P.A.D. II	–	–	–	–
P.A.D. III				
Louisiana	–	–	–	–
Texas and others	–	–	–	–
P.A.D. IV	10,795	1.50	4,321	2.45
P.A.D. V				
California	386,805	53.73	89,254	50.66
Other States	26,597	3.69	12,146	6.89
Total domestic	424,197	58.92	105,721	60.01
Africa				
Algeria	–	–	–	–
Angola	–	–	–	–
Egypt	–	–	–	–
Libya	–	–	–	–
Nigeria	–	–	–	–
Tunisia	–	–	–	–
Subtotal Africa	–	–	–	–
Middle East				
Iran	13,744	1.91	5,920	3.36
Iraq	1,020	.14	515	.29
Israel	–	–	–	–
Kuwait	1,698	.24	–	–
Qatar	1,100	.15	2	0
Saudi Arabia	84,518	11.74	27,056	15.36
United Arab Emirates	11,190	1.55	3,927	2.23
Subtotal Middle East	113,270	15.73	37,420	21.24
South America				
Bolivia	–	–	–	–
Colombia	–	–	–	–
Ecuador	–	–	–	–
Mexico	–	–	–	–
Trinidad	–	–	–	–
Venezuela	8,848	1.23	1,295	.74
Subtotal S. America	25,190	3.50	8,314	4.72
Far East				
Indonesia	68,858	9.56	24,712	14.03
Malaysia	234	.03	–	–
Subtotal Far East	69,092	9.60	24,712	14.03
Canada	88,216	12.25	–	–
Total foreign	295,768	41.08	70,446	39.99
Total crude	719,965	100.00	176,167	100.00

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APPENDIX G

SCALE UP OF CLUSTER RESULTS - DERIVATION OF PRODUCT DEMANDS FOR GRASS ROOTS REFINERIES

A. INTRODUCTION

Appendix F explained how the U.S. refining industry has been simulated by the study of six cluster models, each cluster representing three existing refineries in different regions of the U.S.A. This appendix discusses the method of scale up of the results obtained from the cluster model analysis to represent an aggregate of the total U.S. refining industry. It also describes how the demands for the grassroots refineries were determined.

B. 1973 CALIBRATION SCALE UP

Each cluster model was considered to represent either part of or a complete PAD, with the exception of PAD IV which was not represented by a specific cluster model. The input/output data used in the calibration runs (Appendix I) was then scaled up by making the gasoline production in the cluster model equal to the total gasoline production for each PAD as defined in the BOM annual data for 1973.

PAD II is represented by two cluster models; it has been assumed that the Small Midcontinent cluster represents operations of the Oklahoma/Kansas/Missouri district and that the balance of District II is represented by the Large Midwest cluster.

Similarly in PAD III, it has been assumed that the Louisiana Gulf cluster represents the BOM Louisiana Gulf refining district and the Texas Gulf cluster represents the balance of District III.

Table G-1 gives the input/output data used in the model calibration runs. These data were then scaled up and are compared with the BOM data in Table G-2. For example, for the East Coast cluster a scale up factor

Table G-1. ADL MODEL INPUT/OUTTURN DATA FOR CALIBRATION – 1973

MB/CD

	Louisiana Gulf	Texas Gulf	Large Midwest	Small Midcontinent	East Coast	West Coast
Product Outturns						
Refinery gas/ethane (FOE)	0.32	.84	–	0.54	0.86	0.46
LPG-fuel	5.97	4.96	3.19	1.45	6.73	3.99
LPG-petrochemicals	2.40	3.97	–	0.60	4.13	1.30
Gasoline	118.79	165.62	81.70	37.44	104.93	67.77
Naphtha	0.78	8.80	2.16	0.35	1.28	3.81
BTX	–	6.05	0.94	1.40	1.36	3.90
Jet fuel	18.53	23.49	2.12	0.92	5.76	19.89
Kerosene	5.50	7.70	1.66	0.04	3.39	0.17
Distillate fuel oil	67.80	83.27	40.80	16.04	45.52	22.15
Lube stocks	–	16.49	–	0.34	4.94	0.35
Residual fuel oil	5.35	15.68	7.39	0.23	12.83	30.55
Asphalt	1.55	1.40	3.81	1.81	18.13	2.02
Coke	4.12	4.00	3.68	1.31	–	9.58
Cat cracker feed	2.161	2.00	–	–	–	–
Cat reformer feed	1.164	4.955	–	–	–	–
Total outturns	**234.436**	**349.225**	**147.45**	**62.47**	**209.86**	**165.94**
Inputs						
Isobutane	6.10	2.13	3.70	0.94	0.35	0.50
Normal butane	5.97	2.05	–	0.31	1.76	0.16
Natural gasoline	4.28	16.00	0.93	5.50	5.84	1.30
Natural gas (FOE)	5.40	13.447	0.24	2.046	2.50	6.39
Cat cracker feed	–	–	1.215	.436	11.10	3.597
Cat reformer feed	–	–	.654	.235	5.98	1.937
Crude oil						
Domestic	222.200	294.270	118.265	45.319	43.290	79.381
Foreign	–	37.130	27.280	9.800	144.685	75.816
Total crude	**222.200**	**331.400**	**145.545**	**55.119**	**187.975**	**155.197**
Total inputs	**243.95**	**365.027**	**152.284**	**64.586**	**215.505**	**169.081**

TABLE G-2. COMPARISON OF 1973 B.O.M. DATA AND SCALE UP OF 1973 CALIBRATION INPUT/OUTTURN
(MB/CD)

	PAD I			PAD II									PAD III								
				Ohio, Kans., etc.			Balance PAD			Total PAD			L.A. Gulf			Balance PAD			Total PAD		
	B.O.M. data	A/T	Cluster model scaled up	B.O.M. data	A/T	Cluster model scaled up	B.O.M. data	A/T	Cluster model scaled up	B.O.M. data	A/T	Cluster model scaled up	B.O.M. data	A/T	Cluster model scaled up	B.O.M. data	A/T	Cluster model scaled up	B.O.M. data	A/T	Cluster model scaled up
Inputs																					
Domestic crude	237.3	–	308.5	893.3	–	657.9	1,880.9	–	2,101.5	2,774.2	–	2,759.4	1,617.2	–	1,697.8	3,038.7	–	3,153.1	4,655.9	–	4,850.9
Imported crude	1,264.1	–	1,031.2	28.2	–	142.3	682.5	–	484.7	710.7	–	627.0	46.2	–	–	351.9	–	397.8	398.1	–	397.8
Subtotal	1,501.4	100.0	1,339.7	921.5	60.0	800.2	2,563.4	40.0	2,586.2	3,484.9	100.0	3,386.4	1,663.4	20.0	1,697.8	3,390.6	–	3,550.9	5,054.0	20.0	5,248.7
Isobutane	–	–	2.5	–	–	13.6	–	–	65.7	–	–	79.3	–	–	46.6	–	–	22.8	–	–	69.4
Normal butane	–	–	12.5	–	–	4.5	–	–	–	–	–	4.5	–	–	45.6	–	–	22.0	–	–	67.6
Subtotal	.8	–	15.0	32.3	–	18.1	40.9	–	65.7	73.2	–	83.8	54.4	–	92.2	62.2	–	44.8	116.6	–	137.0
Natural gasoline	.4	–	41.6	32.7	–	78.8	20.7	–	16.5	53.4	–	95.3	66.2	–	32.7	291.2	–	171.4	357.4	–	204.1
Plant condensate	5.8	–	–	.1	–	–	66.9	–	–	67.0	–	–	4.4	–	–	41.2	–	–	45.6	–	–
Unfinished oils	108.2	–	121.7	–	–	9.7	5.1	–	33.2	5.1	–	42.9	41.2	–	–	–	–	–	41.2	–	–
Total	1,616.6	100.0	1,518.0	986.6	60.0	907.8	2,697.0	40.0	2,701.6	3,683.6	100.0	3,609.4	1,829.6	20.0	1,822.7	3,785.2	–	3,767.1	5,614.8	20.0	5,589.8
Purch. natural gas (FOE)	14.0	–	17.8	37.6	–	29.7	19.8	–	4.3	57.4	–	34.0	57.2	–	41.3	264.8	–	144.1	322.0	–	185.4
Outturns																					
Gas/ethane FOE	2.7	–	6.1	2.1	–	7.8	5.8	–	–	7.9	–	7.8	3.6	–	2.4	27.6	–	9.0	31.2	–	11.4
LPG-fuel	39.4	–	48.0	18.2	–	21.0	50.2	–	56.7	68.4	–	77.7	38.9	–	45.6	58.4	–	53.1	97.3	–	98.7
LPG-petrochemicals	17.5	–	29.4	4.0	–	8.7	7.6	–	–	11.6	–	8.7	19.0	–	18.3	44.6	–	42.5	63.6	–	60.8
Gasoline	747.8	–	747.8	543.5	–	543.5	1,451.7	–	1,451.7	1,995.2	–	1,995.2	907.7	–	907.7	1,774.7	–	1,774.7	2,682.4	–	2,682.4
Naphtha	7.0	–	9.1	15.0	10.0	5.1	27.9	–	38.4	42.9	10.0	43.5	17.6	5.0	6.0	100.7	–	94.3	118.3	5.0	100.3
BTX	13.5	–	9.7	6.0	–	20.3	12.0	–	16.7	18.0	–	37.0	1.3	–	–	109.1	–	84.8	110.4	–	84.8
Jet fuel	34.1	–	41.1	32.0	5.0	13.4	113.7	10.0	37.7	145.7	15.0	51.1	140.2	–	141.6	187.8	–	251.7	328.0	–	393.3
Kerosene	19.2	–	34.2	5.7	5.0	.6	48.8	10.0	29.5	54.5	15.0	30.1	55.0	5.0	42.0	79.3	–	82.5	134.3	5.0	124.5
Distillate fuel oil	404.9	50.0	324.4	236.4	–	232.9	587.3	–	725.0	823.7	–	957.9	481.0	–	518.1	878.4	–	892.2	1,359.4	–	1,410.3
Lube stocks	37.0	–	35.2	16.4	–	4.9	16.3	–	–	32.7	–	4.9	24.2	–	–	94.1	–	176.7	118.3	–	176.7
Residual fuel oil	143.2	50.0	91.4	21.1	20.0	3.3	173.7	10.0	131.3	194.8	30.0	134.6	65.1	–	40.9	177.3	–	168.0	242.4	–	208.9
Asphalt	101.7	–	128.2	44.6	20.0	28.3	124.6	10.0	67.7	169.1	30.0	96.0	40.4	10.0	11.8	73.3	–	15.0	113.7	10.0	26.8
Coke-market	15.3	–	–	14.9	–	19.0	36.6	–	65.4	51.5	–	84.4	16.9	–	31.5	32.4	–	42.9	49.3	–	74.4
Unfinished oils	–	–	–	4.5	–	–	–	–	–	4.5	–	–	–	–	25.4	61.8	–	74.5	61.8	–	99.9
Total	1,583.3	100.0	1,495.6	965.3	60.0	908.8	2,656.2	40.0	2,620.1	3,621.5	100.0	3,528.9	1,810.9	20.0	1,791.3	3,699.5	–	3,741.9	5,510.4	20.0	5,533.2
Crude capacity (MB/CD)	1,673.52	–	–	1,000.44	–	–	2,888.972	–	–	3,889.412	–	–	1,893.05	–	–	4,039.876	–	–	5,932.926	–	–
@ 90%	1,506.17	–	–	900.40	–	–	2,600.07	–	–	3,500.47	–	–	1,703.75	–	–	3,635.89	–	–	5,339.63	–	–
Scale up factor	7.127			14.817			17.789						7.641			10.715					

TABLE G 2 (Continued) COMPARISON OF 1973 B.O.M. DATA AND SCALE UP OF 1973 CALIBRATION INPUT/OUTTURN

(MB/CD)

	PAD IV	PAD I-IV				PAD V		Total U.S.			
	B.O.M. A/T	Total cluster model scaled up	Total A/T	Subtotal	Total B.O.M.	B.O.M. data	Cluster model scaled up	Total cluster model scaled up	Total A/T	Subtotal	Total B.O.M.
Intakes											
Domestic crude	371.0	7,918.8	–	–	8,038.4	1,166.9	1,076.0	8,994.8	–	–	9,205.3
Imported crude	44.1	2,056.0	–	–	2,417.0	808.4	1,027.7	3,083.7	–	–	3,225.4
Subtotal	415.1	9,974.8	635.1	10,609.9	10,455.4	1,975.3	2,103.7	12,078.5	635.1	12,713.6	12,430.7
Isobutane	–	151.2	–	–	–	–	6.8	158.0	–	–	–
Normal butane	–	84.6	–	–	–	–	2.2	86.8	–	–	–
Subtotal	9.4	235.8	9.4	245.2	200.0	19.8	9.0	244.8	9.4	254.2	219.8
Natural gasoline	4.6	342.0	4.6	346.6	415.8	23.4	17.6	359.6	4.6	364.2	439.2
Condensate	27.7	–	27.7	27.7	146.1	9.8	–	–	27.7	27.7	155.9
Unfinished oils	.2	64.7	.2	64.9	88.4	37.0	75.0	139.7	.2	139.9	125.4
Total	457.0	10,617.3	677.0	11,294.3	11,305.7	2,065.3	2,205.3	12,822.6	677.0	13,499.6	13,371.0
Purch. natural gas (FOE)	11.9	237.2	11.9	249.1	405.3	61.6	86.6	323.8	11.9	335.7	466.9
Outturn											
Gas/ethane-FOE	.4	25.3	.4	25.7	42.2	3.0	6.2	31.5	.4	31.9	45.2
LPG-fuel	6.0	224.4	6.0	230.4	212.2	33.4	54.1	278.5	6.0	284.5	245.6
LPG-petrochemicals	.2	98.9	.2	99.1	92.9	11.3	17.6	116.5	.2	116.7	104.2
Gasoline	228.2	5,425.4	228.2	5,653.6	5,653.6	918.6	918.6	6,344.0	228.2	6,572.2	6,572.2
Naphtha	8.0	152.9	23.0	175.9	176.2	58.5	51.6	204.5	23.0	227.5	234.7
BTX	.1	111.5	.1	111.6	142.0	14.7	52.9	164.4	.1	164.5	156.7
Jet fuel	14.5	485.5	29.5	515.0	522.3	192.4	269.6	755.1	29.5	784.6	714.7
Kerosene	6.0	178.8	26.0	204.8	214.0	3.6	2.3	181.1	26.0	207.1	217.6
Distillate fuel oil	116.1	2,692.6	165.1	2,857.7	2,703.1	289.7	300.2	2,992.8	165.1	3,157.9	2,992.8
Lube stocks	1.3	216.8	1.3	218.1	189.3	17.6	4.7	221.5	1.3	222.8	206.9
Residual fuel oil	27.0	434.9	107.0	541.9	607.4	364.1	414.1	849.0	107.0	956.0	971.5
Asphalt	30.5	250.0	70.5	320.5	415.0	64.9	27.4	277.4	70.5	347.9	479.9
Coke-market	3.9	158.8	3.9	162.7	120.0	65.4	129.9	288.7	3.9	292.6	185.4
Unfinished oils	–	–	–	–	–	–	–	–	–	–	–
Total	441.2	10,455.8	661.2	11,117.0	11,090.2	2,037.2	2,249.2	12,705.0	661.2	13,366.2	13,127.4
Crude Capacity (MB/CD)	505.721	–	–	–	12,001.579	2,218.737	–	–	–	–	14,220.316
@ 90%	455.15	–	–	–	10,801.42	1,996.86	–	–	–	–	12,798.28
Scale up factor						13.555					

of 7.127 has been used since this is the ratio of the gasoline production of District I to the gasoline production of the East Coast cluster. Table G-2 also contains a middle column entitled "A/T", which stands for atypical configuration. This column is an estimate of the effect of those refineries which do not produce as much motor gasoline on the total output of each district (except PAD IV). By adding the inputs and outputs from these atypical refineries to the scaled up cluster model inputs and outputs, the total cluster model simulation is obtained, which should be comparable to the BOM data. The basis for determining the volume and product mix of the atypical refineries in each region is now discussed.

For PAD I, total product outturns from the cluster model were 1,495.6 MB/CD, while the BOM data indicated 1,583.3; therefore 100 MB/CD has been accounted for via the A/T configuration. The choice of 50 MB/CD as distillate fuel oil and 50 MB/CD as residual fuel oil was made because these products were in short supply from the cluster model; also, the atypical refineries in PAD I produce heavy products predominantly. It should be noted that the crude supply is equal to the products (i.e. no processing loss or gain). However, since the A/T crude intake has only a minor effect on crude slate, this is not important.

In PAD II, which is represented by two cluster models, the addition of 60 MB/CD A/T configuration for the Small Midcontinent and 40 MB/CD for the Large Midwest brought the product outturns from the cluster models close to the BOM statistics.

In PAD III, only a nominal 20 MB/CD of A/T configuration was used to balance the entire district. PAD IV is not simulated by a cluster model and therefore the basic BOM data is by definition equal to the A/T configuration needed to balance the district.

The model simulation for PAD's I-IV is obtained by combining the total cluster model output with the total A/T (including PAD IV), for comparison with the BOM data for Districts I-IV. Of course, as shown in Table G-2, the gasoline productions are equal. Other products check quite well with the cluster models being approximately 150 MB/CD high on distillate fuel and slightly more than 100 MB/CD low on total residual products (residual fuel, plus asphalt, plus coke). Total product outturns differed by less than

30 MB/CD and total intakes by about 10 MB/CD. The scaled up model runs were about 150 MB/CD high on crude and 50 MB/CD on butanes, but these were offset by natural gasoline and condensate.

The scale up for PAD V is presented next. After scaling up the gasoline the results show a greater production of other products (primarily jet fuel, residual fuel oil and coke) than the BOM data. Thus, it was concluded that no A/T configuration would be added for this region.

C. DERIVATION OF MODEL FIXED INPUTS AND OUTPUTS FOR FUTURE YEARS

The crude oils and other fixed inputs used in the model runs for the years 1977, 1980 and 1985 were based on the inputs used in the calibration runs with certain modifications, as shown in Tables G-3, G-4, and G-5.

The choice of crude oil types has already been discussed in Appendix A. The amount of crude oil processed was kept constant in all three future years studied, to simulate no expansion of these refineries (expansion was included in the grassroots models).

Butanes and natural gasoline inputs were reduced from the calibration levels to reflect a gradual reduction in availability of these materials. The volumes used in the calibration runs were reduced by 10% in 1977, 20% in 1980 and 30% in 1985.

Natural gas purchased by the cluster model refineries was completely phased out by 1985 to reflect our forecast of the declining production of natural gas. In 1977, 75% of the volumes purchased in the calibration runs were used and in 1980 the volumes purchased were 50% of those in calibration.

Product demands with the exception of LPG and gasoline were fixed, based on the demands used in the calibration run. The actual demands used in the cluster model runs for future years were the calibration demands ratioed up or down based on the total fixed input to the cluster model (crude oil, butanes, natural gasoline, natural gas and unfinished oil). For example, the total inputs to the Texas Gulf cluster model in the calibration run amounted to 365,027 barrels per calendar day (331,400 of crude oil, 4,180 of butanes, 16,000 of natural gasoline and 13,447 of natural gas). In the 1985 model runs the total inputs to the Texas Gulf

Table G-3. L.P. MODEL INPUT/OUTTURNS 1977

(MB/CD)

	East Coast	Large Midwest.	Small Midcontinent	Louisiana Gulf	Texas Gulf	West Coast
Fixed intakes						
Domestic crude	–	104.575	42.744	217.993	291.633	87.349
Imported crude	197.917	38.875	12.197	–	36.782	76.840
Subtotal	**197.917**	**143.450**	**54.941**	**217.993**	**328.415**	**164.189**
Isobutane	0.315	3.330	0.846	5.490	1.917	0.450
Normal butane	1.584	–	0.279	5.373	1.845	0.144
Natural gasoline	5.840	0.837	4.950	3.852	14.400	1.170
Natural gas (FOE)	1.875	0.180	1.535	4.050	10.085	4.793
Unfinished oils	17.060	1.869	0.671	–	–	5.534
Total	**224.611**	**149.666**	**63.222**	**236.758**	**356.662**	**176.280**
Fixed outturns						
Gas/ethane (FOE)	0.896	–	0.528	0.311	0.821	0.480
LPG - petrochem.	4.303	–	0.587	2.329	3.879	1.357
Naphtha	1.333	2.122	0.343	0.757	8.598	3.978
BTX	1.417	0.924	1.370	–	5.911	4.072
Jet fuel	6.001	2.083	0.900	17.984	22.952	20.765
Kerosene	3.532	1.631	0.039	5.338	7.524	0.177
Distillate	47.431	40.090	15.698	65.801	61.363	23.125
Lube stocks	5.147	–	0.333	–	16.112	0.365
Resid. fuel oil	13.368	7.261	0.225	5.192	15.321	31.894
Asphalt	18.891	3.744	1.771	1.504	1.368	2.109
Coke	––	3.616	1.282	3.999	3.908	10.002
Unfinished oils	–	–	–	3.198	6.796	–
Total	**102.319**	**61.471**	**23.076**	**106.413**	**174.553**	**98.324**
Variable outturns						
Scenario **Product**						
A Gasoline	110.972	79.463	36.922	119.104	164.006	71.380
LPG	4.634	2.716	1.320	3.223	5.263	3.458
B Gasoline	110.424	79.308	36.889	119.328	163.300	71.208
LPG	4.402	2.793	1.331	2.843	5.849	3.623
C Gasoline	109.109	79.295	36.732	118.346	161.267	69.634
LPG	5.579	3.121	1.454	3.588	7.303	4.033
D Gasoline	110.236	78.838	36.618	118.308	161.267	69.631
LPG	5.246	3.407	1.601	3.623	7.303	4.035
E Gasoline	109.771	78.862	36.579	118.369	161.267	69.627
LPG	5.883	3.392	1.648	3.483	7.303	4.038
F Gasoline	108.987	77.040	35.541	116.558	161.288	69.609
LPG	5.140	3.415	1.552	3.831	6.919	3.997

Table G-4. L.P. MODEL INPUT/OUTTURNS 1980
(MB/CD)

Fixed intakes		East Coast	Large Midwest	Small Midcontinent	Louisiana Gulf	Texas Gulf	West Coast
Domestic crude		–	92.812	39.447	217.993	291.633	164.190
Imported crude		197.910	50.638	15.494	–	36.782	–
Subtotal		197.910	143.450	54.941	217.993	328.415	164.190
Isobutane		0.280	2.960	0.752	4.880	1.704	0.400
Normal butane		1.400	–	0.248	4.776	1.640	0.128
Natural gasoline		5.840	0.744	4.400	3.424	12.800	1.040
Natural gas (FOE)		1.250	0.120	1.023	2.700	6.724	3.195
Unfinished oils		17.080	1.869	0.671	–	–	5.534
Total		223.760	149.143	62.035	233.773	351.283	174.487
Fixed outturns							
Gas/ethane (FOE)		0.892	–	0.518	0.307	0.808	0.475
LPG-petrochem.		[illegible].286	–	0.576	2.300	3.820	1.343
Naphtha		1.328	2.115	0.336	0.747	8.468	3.936
BTX		1.411	0.920	1.344	–	5.822	4.029
Jet Fuel		5.978	2.076	0.883	17.757	22.604	20.548
Kerosene		3.518	1.625	0.038	5.271	7.410	0.176
Distillate		47.249	39.948	15.400	64.972	80.131	22.883
Lube stocks		5.127	–	0.326	–	15.868	0.362
Resid. fuel oil		13.317	7.236	0.221	5.127	15.089	31.561
Asphalt		18.818	3.730	1.738	1.485	1.347	2.087
Coke		–	3.603	1.258	3.948	3.849	9.897
Unfinished oils		–	–	–	3.198	6.693	–
Total		101.924	61.253	22.638	105.112	171.909	97.297
Variable outturns							
Scenario	**Product**						
A	Gasoline	111.524	78.635	36.219	118.130	161.364	71.735
	LPG	3.954	2.807	1.261	2.855	5.268	2.989
B	Gasoline	109.817	78.070	35.874	117.446	159.451	70.973
	LPG	4.910	3.498	1.569	3.324	6.950	3.732
C	Gasoline	108.723	77.425	35.472	116.110	158.202	70.398
	LPG	5.561	3.999	1.948	4.315	7.516	3.016
D	Gasoline	109.435	77.024	35.437	115.505	158.062	69.814
	LPG	5.568	4.060	1.981	4.541	7.600	3.723
E	Gasoline	108.862	76.906	35.302	114.842	157.874	69.707
	LPG	6.543	4.100	1.957	4.768	7.615	3.740

Table G-5. L.P. MODEL INPUT/OUTTURNS-1985

(MB/CD)

		East Coast	Large Midwest	Small Midcontinent	Louisiana Gulf	Texas Gulf	West Coast
Fixed intakes							
Domestic crude		–	85.353	36.151	217.993	291.633	164.190
Imported crude		197.915	58.097	18.790	–	36.782	–
Subtotal		**197.915**	**143.450**	**54.941**	**217.993**	**328.415**	**164.190**
Isobutane		.245	2.590	.658	4.700	1.491	.350
Normal butane		1.232	–	.217	4.179	1.435	.112
Natural gasoline		5.840	.651	3.850	3.000	11.200	.910
Unfinished oils		17.080	1.869	.671	–	–	5.534
Total		**222.312**	**148.560**	**60.337**	**229.872**	**342.541**	**171.096**
Fixed outturns							
Gas/ethane (FOE)		.887	–	.504	.301	.788	.470
LPG-petrochem.		4.260	–	.560	2.257	3.725	1.331
Naphtha		1.320	2.107	.327	.734	8.258	3.900
BTX		1.403	.917	1.307	–	5.677	3.993
Jet fuel		5.941	2.068	.859	17.427	22.043	20.364
Kerosene		3.497	1.619	.037	5.173	7.226	.174
Distillate fuel oil		46.954	39.790	14.973	63.763	78.141	22.678
Lube stocks		5.096	–	.317	–	15.474	.358
Residual fuel oil		13.234	7.207	.215	5.032	14.714	31.279
Asphalt		18.700	3.716	1.690	1.458	1.314	2.068
Coke market		–	3.589	1.223	3.870	3.764	7.790
Unfinished oils		–	–	–	3.198	6.527	–
Total		**101.292**	**61.013**	**22.012**	**103.213**	**167.641**	**94.405**
Variable outturns							
Scenario	**Product**						
A	Gasoline	110.785	78.570	35.174	116.277	157.251	71.613
	LPG	3.981	2.683	1.368	2.878	5.392	–
B/C	Gasoline	106.915	76.615	34.016	112.488	152.330	70.290
	LPG	6.119	4.063	1.720	5.082	8.014	0.044
D	Gasoline	107.116	75.028	33.045	111.981	152.720	68.171
	LPG	6.460	4.542	1.689	5.669	7.860	3.288
E	Gasoline	104.981	74.078	33.033	102.790	142.134	68.350
	LPG	9.221	5.423	1.655	14.309	15.997	3.054
F	Gasoline	107.447	74.830	32.775	113.770	152.850	69.730
	LPG	5.592	4.238	1.622	0.910	7.667	–

cluster model totaled 342,541 barrels per calendar day (328,415 of crude oil, 2,926 of butanes and 11,200 of natural gasoline). The product demands for the Texas Gulf cluster for 1985 were derived by multiplying the calibration run demands by a factor of 0.9384 (the ratio of total input in 1985 to the total input in the calibration run).

Also listed in Tables G-3, G-4 and G-5 are the variable outturns of LPG and gasoline that were produced in each scenario studied.

D. SCALE UP OF RESULTS FOR FUTURE YEARS

The model results for the study years of 1977, 1980 and 1985 were scaled up using the atypical concept derived from the calibration results. In 1977, scale up factors were derived as in the calibration scale up, using total gasoline demand. In 1980 and 1985 the scale up factors used, however, were based on total crude run in each cluster and the effective crude oil distillation capacity for the region simulated by that cluster. The scale up factors used were calculated by making the crude run in each region equal to the effective crude oil distillation capacity for that region. Effective crude oil distillation capacity was defined as 90% of the 1973 calendar day rated capacity, which is similar to historical capacity utilization. This capacity is shown for each region in Table G-2.

1. 1977 Scale Up

The scale up of results for 1977 was based on meeting the gasoline demand for the total U.S. from scaled up cluster model gasoline productions, atypicals and imports only. Since crude capacity utilization was less than 90%, grassroots refineries were not needed in 1977.

This scale up method results in a different scale up factor for each scenario, since the cluster models produce different gasoline volumes in each scenario and each scenario is scaled up to the same total U.S. demand for gasoline. Therefore, in 1977 the penalties for meeting the proposed regulations will be based on the loss of other products in addition to additional crude runs required while continuing to produce the same gasoline volume. Table G-3 gives the fixed inputs and outputs for the 1977 cluster runs and the variable gasoline and LPG productions for each scenario. Table G-6 gives the scaled up fixed inputs, fixed outputs and LPG for each scenario when producing the same volume of gasoline.

Table G-6 SCALE UP INPUT/OUTTURNS 1977

Scenario	Input/outturn	East Coast		Large Midwest		Small Midcont.		Louisiana Gulf		Texas Gulf		West Coast	
		Factor	MB/CD	Factor	MB/CD	Factor	MB/CD	Factor	MB/CD	Factor	MB/CD	Factor	MB/CD
A	Fixed input	7.649	1,718.0	18.217	2,726.5	16.417	1,037.9	7.856	1,856.0	11.127	3,968.6	13.392	2,360.7
	Fixed output		782.6		1,119.8		378.8		836.0		1,942.3		1,316.8
	LPG		35.4		49.5		21.7		25.3		58.6		46.3
B	Fixed input	7.666	1,721.9	18.258	2,732.6	16.454	1,040.3	7.874	1,864.2	11.152	3,977.5	13.423	2,366.2
	Fixed output		784.4		1,122.3		379.7		837.9		1,946.6		1,319.8
	LPG		33.7		51.0		21.9		22.4		65.2		48.6
C	Fixed input	7.713	1,732.4	18.369	2,749.2	16.609	1,050.1	7.921	1,875.4	11.220	4,001.7	13.727	2,419.8
	Fixed output		789.2		1,129.2		383.3		842.9		1,958.5		1,349.7
	LPG		43.0		57.3		24.1		28.4		81.9		55.4
D	Fixed input	7.715	1,732.9	18.376	2,750.3	16.615	1,050.4	7.924	1,876.1	11.224	4,003.2	13.728	2,420.0
	Fixed output		789.4		1,129.6		383.4		843.2		1,959.2		1,349.8
	LPG		40.8		62.6		26.6		28.7		82.0		55.4
E	Fixed input	7.720	1,734.0	18.387	2,751.9	16.625	1,051.1	7.929	1,877.3	11.231	4,005.7	13.729	2,420.1
	Fixed output		789.9		1,130.3		383.6		843.7		1,960.4		1,349.9
	LPG		40.8		62.6		27.4		27.6		82.0		55.4
F	Fixed input	7.766	1,744.3	18.450	2,761.3	16.725	1,057.4	7.977	1,888.6	11.299	4,029.9	13.732	2,420.7
	Fixed output		794.6		1,134.1		385.9		848.9		1,972.3		1,350.2
	LPG		39.9		58.7		26.0		30.6		78.2		54.9

In evaluating the penalties associated with any regulation, the fixed inputs and outputs have both been considered as crude oil.

Product imports into the U.S.A. have been assumed to continue at similar levels as experienced in 1973 in all the study years. Table G-7 gives the assumptions of atypical refinery inputs and outputs for 1977, 1980 and 1985. The atypical refinery data for PAD IV were based on the assumption that total crude (plus condensate) was 90% of the rated calendar day capacity for this district. Product outputs were then ratioed on the basis of total output in each year compared with the calibration results. The data for atypical refineries for the remaining PAD districts were based on a 2% per annum escalation of the 1973 volumes, but assuming zero growth from 1973 to 1975.

2. 1985 Scale Up

The cluster model input/output data for Scenarios A, B/C, D, E and F is given in Table G-5. All input data and output data for each cluster are the same for all scenarios with the exception of gasoline and LPG which were allowed to vary from scenario to scenario.

The scale up of the cluster results is given in Table G-8. For example, for the East Coast cluster, the amount of crude run in the L.P. model was 197.915 MB/CD. The effective crude oil distillation capacity for this region was 1506.17 MB/CD and therefore a scale up factor of 7.610 was used (the ratio of effective capacity over model crude run). The atypical and import volumes are then added to the scaled up cluster volumes to give total supply of products. These are then compared with the forecast demand for the major product groups (Gasoline, Jet Fuel, Kerosene, Distillate Heating Oil and Residual Fuel Oil) in order to derive the demand for grassroots refining. For example, in Table G-8 the total supply of jet fuel from existing refineries in Districts I-IV in 1985 is 634,600 barrels per calendar day. The forecasted demand for all refineries (existing plus additional capacity built by 1985) is 780,000 barrels per calendar day. Therefore, the jet fuel production required of the grassroots refineries in Districts I-IV was set at 145,400 barrels per calendar day. The grassroots productions of kerosene, distillate fuel oil and residual fuel oil were determined in a similar manner. The grassroots

Table G-7. ATYPICAL REFINERY INTAKE/OUTTURN SUMMARY

(MB/CD)

	1973		1977		1980		1985	
	PAD IV	Other[a]	PAD IV	Other[a]	PAD IV	Other[a]	PAD IV	Other[a]
Crude (+cond.)	442.8	220.0	455.2	228.9	455.2	242.9	455.2	268.2
C4's	9.4	–	8.5	–	7.9	–	7.0	–
Natural gasoline	4.6	–	4.2	–	4.1	–	3.6	–
Unfinished	0.2	–	0.2	–	0.2	–	0.2	–
Natural gas (FOE)	11.9	–	8.9	–	6.0	–	.0	–
Total	468.9	220.0	477.0	228.9	473.4	242.9	466.0	268.2
Gas	0.4	–	0.4	–	0.4	–	0.4	–
LPG-fuel	6.0	–	6.1	–	6.1	–	6.0	–
LPG-petrochemical	0.2	–	0.2	–	0.2	–	0.2	–
Gasoline	228.2	–	232.1	–	230.4	–	226.8	–
Naphtha	8.0	15.0	8.1	15.6	8.1	16.6	8.0	18.3
BTX	0.1	–	0.1	–	0.1	–	0.1	–
Jet fuel	14.5	15.0	14.8	15.6	14.6	16.6	14.4	18.3
Kerosene	6.0	20.0	6.1	20.8	6.1	22.1	6.0	24.4
Distillate	115.1	50.0	117.1	52.1	116.2	55.2	114.4	61.0
Lubes	1.3	–	1.3	–	1.3	–	1.3	–
Residual	27.0	80.0	27.5	83.2	27.2	88.3	26.8	97.5
Asphalt	30.5	40.0	31.0	51.6	30.8	44.1	30.3	48.8
Coke	3.9	–	4.0	–	3.9	–	3.9	–
Total	441.2	220.0	448.8	228.9	445.4	242.9	438.6	268.3

[a] Total of PADS I, II, III and V.

Table G-8. SCALE UP INPUT/OUTPUT – 1985

(MB/CD)

		PAD Districts I-IV											PAD District V					Total U.S.				
		East Coast	Large Midwest	Small Midcont.	Louis. Gulf	Texas Gulf	Subtotal I-IV	A/T	Major product imports	Total intake/ supply	Major product demand	Grass Roots required outturn	Cluster scale up	Major product imports	Total supply	Major product demand	Grass Roots required outturn	Cluster scale up + A/T	Major product imports	Total supply	Major product demand	Grass Roots required outturn
Fixed intakes																						
Domestic crude		–	1,547.0	592.4	1,703.8	3,228.7	7,071.9						1,996.9									
Imported crude		1,506.1	1,053.0	307.9	–	407.2	3,274.2						–									
Subtotal		1,506.1	2,600.0	900.3	1,703.8	3,635.9	10,346.1	723.4		11,069.5			1,996.9					13,066.4				
Isobutane		1.9	46.9	10.8	36.7	16.5	112.8						4.3									
Normal butane		9.4	–	3.6	32.7	15.9	61.6						1.4									
Subtotal		11.3	46.9	14.4	69.4	32.4	174.4	7.0		181.4			5.7					187.1				
Natural gasoline		44.4	11.8	63.1	23.4	124.0	266.7	3.6		270.3			11.1					281.4				
Unfinished oils		130.0	32.9	11.0	–	–	77.6	.2		77.8			67.3					145.1				
Total		1,691.8	2,691.6	988.8	1,796.6	3,792.3	10,864.8	734.2		11,599.0			2,081.0					13,680.0				
Fixed outturns																						
Gas/ethane-FOE		6.8	–	8.3	2.4	8.7	25.2	.4		25.6			5.7		5.7			32.3		32.3		
LPG-petrochemicals		32.4	–	9.2	17.6	41.2	100.4	.2		100.6			16.2		16.2			116.8		116.8		
Naphtha		10.0	38.2	5.4	5.7	91.4	150.7	26.3		177.0			47.4		47.4			224.4		224.4		
BTX		10.7	16.8	21.4	–	62.9	111.6	.1		111.7			48.6		48.6			160.3		160.3		
Jet fuel		45.2	37.5	14.1	136.2	244.0	477.0	32.7	124.9	634.6	780.0	145.4	247.7	58.4	306.1	399.8	93.7	757.4	183.3	940.7	1,179.8	239.1
Kerosene		26.6	29.3	.6	40.4	80.0	176.9	30.4	2.2	209.5	252.3	42.8	2.1		2.1			209.4	2.2	211.6	252.3	42.8
Distillate fuel oil		357.3	721.2	245.4	498.4	865.1	2,687.4	175.4	381.0	3,243.8	3,948.0	704.2	275.8	11.2	287.0	379.6	92.6	3,138.6	392.2	3,530.8	4,327.6	796.8
Lube stocks		38.8	–	5.2	–	171.3	215.3	1.3		216.6			4.4		4.4			221.0		221.0		
Residual fuel oil		100.7	130.6	3.5	39.3	162.9	437.0	124.3	1,797.7	2,359.0	2,852.0	493.0	380.4	55.0	435.4	571.8	136.4	941.7	1,852.7	2,794.4	3,423.8	629.4
Asphalt		142.3	67.4	27.7	11.4	14.5	263.3	79.1		342.4			25.2		25.2			367.6		367.6		
Coke-market		–	65.1	20.0	30.2	41.6	156.9	3.9		160.8			94.7		94.7			255.5		255.5		
Unfinished oils		–	–	–	25.0	72.3	–	–		–			–		–			–		–		
Total		770.8	1,105.9	360.8	806.6	1,855.9	4,802.7	474.1	2,305.8	7,582.6	7,832.3	1,385.4	1,148.2	124.6	1,272.8	1,351.2	322.7	6,425.0	2,430.4	8,855.4	9,183.5	1,708.1
Variable outturns																						
Scenario	Product																					
A	Gasoline	843.1	1,420.1	576.4	908.8	1,740.9	5,489.3	226.8	130.2	5,846.3	7,050.3	1,204.0	871.0	3.4	874.4	1,123.9	249.5	6,587.1	133.6	6,720.7	8,174.2	1,453.5
	LPG-fuel	30.3	49.9	22.4	22.5	59.7	184.8	6.0		190.8			–		–			190.8		190.8		
	All products	1,644.2	2,575.9	959.6	1,737.9	3,656.5	10,476.8	706.9	2,436.0	13,619.7			2,019.2	128.0	2,147.2	2,475.1	572.2	13,202.9	2,564.0	15,766.9		
B/C	Gasoline	813.6	1,388.6	557.5	879.2	1,686.4	5,325.3	226.8	130.2	5,682.3	7,050.3	1,368.0	854.9	3.4	858.3	1,123.9	265.6	6,407.0	133.6	6,540.6	8,174.2	1,633.6
	LPG-fuel	46.6	73.6	28.2	39.7	88.7	276.8	6.0		282.8			0.5		0.5			283.3		283.3		
	All products	1,631.0	2,588.1	946.5	1,725.5	3,631.0	10,404.8	706.9	2,436.0	13,547.7			2,003.6	128.0	2,131.6	2,475.1	588.3	13,115.3	2,564.0	15,679.3		
D	Gasoline	815.2	1,359.9	541.5	875.2	1,690.8	5,282.6	226.8	130.2	5,639.6	7,050.3	1,410.7	829.1	3.4	832.5	1,123.9	291.4	6,338.5	133.6	6,472.1	8,174.2	1,702.1
	LPG-fuel	49.2	82.3	27.7	44.3	87.0	290.5	6.0		296.5			40.0		40.0			336.5		336.5		
	All products	1,635.2	2,548.1	930.0	1,726.1	3,633.7	10,375.8	706.9	2,436.0	13,518.7			2,017.3	128.0	2,145.3	2,475.1	571.8	13,100.0	2,564.0	15,664.0		
E	Gasoline	798.9	1,342.7	541.3	803.4	1,573.6	5,059.9	226.8	130.2	5,416.9	7,050.3	1,633.4	831.3	3.4	834.7	1,123.9	289.2	6,118.0	133.6	6,251.6	8,174.2	1,922.6
	LPG-fuel	70.2	98.3	27.1	111.8	177.1	484.5	6.0		490.5			37.1		37.1			527.6		527.6		
	All products	1,639.9	2,546.9	929.2	1,721.8	3,606.6	10,347.1	706.9	2,436.0	13,490.0			2,016.6	128.0	2,144.6	2,475.1	611.9	13,070.6	2,564.0	15,634.6		
F	Gasoline	817.7	1,356.3	537.1	889.2	1,682.2	5,292.5	226.8	130.2	5,649.5	7,050.3	1,400.8	848.1	3.4	851.5	1,123.9	272.4	6,367.4	133.6	6,501.0	8,174.2	1,673.2
	LPG-fuel	42.6	76.8	26.6	7.1	84.9	238.0	6.0		244.0			–		–			244.0		244.0		
	All products	1,631.1	2,539.0	924.5	1,702.9	3,633.0	10,333.2	706.9	2,436.0	13,476.1			1,996.3	128.0	2,124.3	2,475.1	595.1	13,036.4	2,564.0	15,600.4		
Scale up factor		7.610	18.125	16.388	7.816	11.071	–	–	–	–	–	–	12.162	–	–	–	–	–	–	–	–	–

production of gasoline varied from scenario to scenario because of the effects of the proposed regulations but was also determined the same way.

The results of the scale up exercise indicated that 15 "new refineries" (these can be new plants or additions to existing ones) with a capacity of approximately 200,000 barrels per calendar day will be required to meet East of the Rockies product demands by 1985. Three "new refineries' will be required to meet West of the Rockies product demands by 1985. In determining the L.P. model inputs for the grassroots cases, the grassroots volumes shown in Table G-8 were therefore divided by 15 for the East of the Rockies model and by 3 for the West of the Rockies model.

The forecast product demand was derived by using a "simulated" product demand pattern obtained by using the scaled up 1973 calibration output for each cluster model combined with the atypicals and the 1973 imports (see Appendix B). This was done to prevent minor discontinuities being leveraged to result in unreasonable grassroots requirements.

This method of determining the demand for grassroots refining does not consider the increased demand for specialty products and therefore the total demand shown for grassroots refining in 1985 will be deficient by the increased demand for specialty products. By comparison with Appendix B, the simulated 1985 product demands are 92% of our forecast of total product demand. The results for the grassroots refinery model runs have therefore been scaled up by a further factor of 1.087 (1 divided by .92) to reflect the need to meet total product demand.

3. 1980 Scale Up

The scale up for 1980 was done in an identical pattern to that for 1985. The scale up results are shown in Table G-9. The results showed that 6 to 7 new grassroots refineries of approximately 200 MB/CD would be required to meet the product demand of PAD I-IV by 1980, included in the total of 15 refineries by 1985 discussed above. Again, these "new refineries" include both major expansions of existing refineries and grassroots refineries.

Two new refineries would be required to meet PAD V product demands by 1980, included in the total of 3 refineries by 1985 discussed above.

TABLE G-9. SCALE UP INPUT/OUTPUT – 1980

(MB/CD)

	PAD DISTRICTS I-IV											PAD DISTRICT V					TOTAL U.S.				
		East Coast	Large Midwest	Small Midcont.	Louis. Gulf	Texas Gulf	Subtotal I-IV	A/T	Major product imports	Total intake/supply	Major product demand	Gross roots required outturn	Cluster scale-up	Major product imports	Total supply	Major product demand	Gross roots required outturn	Cluster scale-up + A/T	Major product imports	Total supply	Major product demand	Gross roots required outturn
Fixed intakes																					
Domestic crude		–	1,682.2	646.4	1,703.8	3,228.7	7,261.2						1,988.3									
Imported crude		1,506.1	917.8	253.9	–	407.2	3,084.9						–									
Subtotal		1,506.1	2,600.0	900.3	1,703.8	3,635.9	10,346.1	698.1		11,044.2			1,988.3					13,032.5				
Isobutane		2.1	53.7	12.3	38.1	18.9	125.1						4.8									
Normal butane		10.7	–	4.1	37.3	18.2	70.3						1.6									
Subtotal		12.8	53.7	16.4	75.4	37.1	195.4	7.9		203.3			6.4					209.7				
Natural gasoline		44.4	13.5	72.1	26.8	141.7	298.5	4.1		302.6			12.6					315.2				
Natural gas (FOE)		9.5	2.2	16.8	21.1	74.4	124.0	6.0		130.0			38.7					168.7				
Unfinished oils		130.0	33.9	11.0	–	–	174.9	0.2		175.1			67.0					242.1				
Total		1,702.8	2,703.3	1,016.6	1,827.1	3,889.1	11,138.9	716.3		11,855.2			2,113.0					13,968.2				
Fixed outturns																					
Gas/ethane (FOE)		6.8	–	8.5	2.4	8.9	26.6	0.4		27.0			5.8		5.8			32.8		32.8		
LPG-petrochemicals		32.6	–	9.4	18.0	42.3	102.3	0.2		102.5			16.3		16.3			118.8		118.8		
Naphtha		10.1	38.3	5.5	5.8	93.7	153.4	24.7		178.1			47.7		47.7			225.8		225.8		
BTX		10.7	16.7	22.0	–	64.5	113.9	0.1		114.0			48.8		48.8			162.8		162.8		
Jet Fuel		45.5	37.8	14.5	138.8	250.2	486.8	31.2	124.9	642.7	706.5	63.8	248.8	58.4	307.2	362.1	54.9	766.6	183.3	949.9	1,068.6	118.7
Kerosene		26.8	29.5	0.6	41.2	82.0	180.1	28.2	2.2	210.5	228.5	18.0	2.1	–	2.1			210.4	2.2	212.6	228.5	18.0
Distillate fuel oil		359.6	724.1	252.4	507.8	887.1	2,731.0	171.4	381.0	3,283.4	3,575.8	292.4	277.1	11.2	288.3	343.8	55.5	3,179.5	392.2	3,571.7	3,919.6	347.9
Lube stocks		39.0	–	5.3	–	175.7	220.0	1.3		221.3			4.4		4.4			225.7		225.7		
Residual fuel oil		101.3	131.2	3.6	40.1	167.1	443.3	115.5	1,797.7	2,356.5	2,583.1	226.6	382.2	55.0	437.2	517.9	80.7	941.0	1,852.7	2,793.7	3,101.0	307.3
Asphalt		143.2	67.6	28.5	11.6	14.9	265.8	74.9		340.7			25.3		25.3			366.0		366.0		
Coke-market		–	65.3	20.8	30.9	42.6	159.4	3.9		163.3			119.9		119.9			283.2		283.2		
Unfinished oils		–	–	–	25.0	74.1	99.1	–		99.1			–		–			99.1		99.1		
Total		775.6	1,110.3	370.9	821.6	1,903.1	4,981.5	451.8	2,305.8	7,739.1	7,093.9	600.8	1,178.4	124.6	1,303.0	1,223.8	191.1	6,611.7	2,430.4	9,042.1	8,317.7	791.9
Variable outturns																					
Scenario	Product																				
A	Gasoline	848.7	1,425.3	593.6	923.3	1,786.5	5,577.4	230.4	130.2	5,938.0	6,385.7	447.7	868.7	3.4	872.1	1,018.0	145.9	6,676.5	133.6	6,810.1	7,403.7	593.6
	LPG-fuel	30.1	50.9	20.7	27.3	53.3	182.3	6.1		188.4			36.2		36.2			224.6		224.6		
	All products	1,654.4	2,586.5	985.2	1,767.2	3,747.5	10,741.2	698.3	2,436.0	13,875.5			2,073.3	128.0	2,201.3			13,512.8	2,564.0	16,076.8		
B	Gasoline	835.7	1,415.0	587.9	918.0	1,765.3	5,521.9	230.4	130.2	5,882.5	6,385.7	503.2	858.5	3.4	862.9	1,018.0	156.1	6,611.8	133.6	6,745.4	7,403.7	658.3
	LPG-fuel	37.4	63.4	25.7	26.0	76.9	229.4	6.1		235.5			45.2		45.2			280.7		280.7		
	All products	1,648.7	2,588.7	984.5	1,765.6	3,746.3	10,732.8	698.3	2,436.0	13,867.1			2,073.1	128.0	2,201.1			13,504.2	2,564.0	16,068.2		
C	Gasoline	827.4	1,403.3	581.3	907.5	1,751.5	5,471.0	230.4	130.2	5,831.6	6,385.7	554.1	852.5	3.4	855.9	1,018.0	162.1	6,553.9	133.6	6,687.5	7,403.7	716.2
	LPG-fuel	47.3	72.8	31.9	33.7	83.2	263.6	6.1		269.7			38.5		38.5			308.2		308.2		
	All products	1,645.3	2,586.1	984.1	1,762.8	3,737.8	10,706.1	698.3	2,436.0	13,840.4			2,067.4	128.0	2,195.4			13,471.8	2,564.0	16,035.8		
D	Gasoline	832.8	1,386.1	580.7	902.6	1,749.9	5,462.3	230.4	130.2	5,822.9	6,385.7	562.8	845.4	3.4	848.8	1,018.0	169.2	6,538.1	133.6	6,671.7	7,403.7	732.0
	LPG-fuel	42.4	73.6	32.5	35.5	84.1	268.1	6.1		274.2			45.1		45.1			319.3		319.3		
	All products	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]			[illegible]	128.0	[illegible]			[illegible]	2,564.0	[illegible]		
E	Gasoline	838.4	1,393.9	578.5	887.6	1,747.8	5,446.2	230.4	130.2	5,806.8	6,385.7	578.9	844.2	3.4	847.6	1,018.0	170.4	6,520.8	133.6	6,654.4	7,403.7	749.3
	LPG-fuel	48.8	74.3	32.1	37.3	84.3	277.8	6.1		283.9			45.3		45.3			329.2		329.2		
	All products	1,653.8	2,578.5	981.5	1,756.5	3,728.2	10,695.5	698.3	2,436.0	13,829.8			2,067.9	128.0	2,195.9			13,461.7	2,564.0	16,025.7		
Scale up factor		7.810	18.125	18.348	7.816	11.671							12.110									

E. SCALE UP OF CAPITAL INVESTMENTS

The scale up factors derived in the first part of this Appendix have been used to scale up all the L.P. model results except the capital investment requirements. In determining the capital investment requirements on an aggregate U.S. basis it was felt that the requirements of the small to medium refiner (refineries with capacities below 75,000 barrels per day) would not be adequately reflected if the scale up factors derived in the first part of this Appendix were applied to the cluster model capital investment requirements.

The only cluster model which represented the small to medium refiner was the Small Midcontinent model. Based on the derived scale up factors the investments resulting from this model would carry a weighting factor of 21% of the total scaled up cluster model investments. In fact, refineries with capacities less than 75,000 barrels per day represent some 30% of the total U.S. refining capacity. To correct for the fact that the straightforward scale up method would not fully account for the small refiners and based on an aggregate of the study results obtained, the total cluster model investments (this does not include the grassroots model investment) were scaled up by a further 17%.

This factor of 17% was derived by calculating the investment requirements associated with a particular regulation for each of the cluster models on a dollar per barrel per day basis. Then, the total capital investment penalty for the regulation was calculated by multiplying the dollar per barrel capital requirement of the Small Midcontinent model by the capacity of refineries in PAD I, for example, with a refinery capacity below 75,000 BPD. To this figure was added the dollar per barrel capital requirement for the East Coast model multiplied by the capacity of refineries in PAD I with individual refinery capacities above 75,000 BPD. When this procedure was repeated for all PAD's, the total U.S. capital investment was found to be 17% above that obtained by a direct use of the scale up factors derived earlier in this Appendix. When this procedure was used for the regulations considered in the three companion reports, these deviations were found to range from 15% to 21%. Therefore, a constant value of 17% was used for all the regulations studied.

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APPENDIX H

TECHNICAL DOCUMENTATION

This appendix provides basic yields, blending properties, operating costs and investment data which comprise the ADL Refinery Modeling System. Over the past several years ADL has accumulated and continually updated a data bank containing crude assays & other relevant data for refinery simulation purposes. The sources for this basic data are many and varied, including technical literature, data received from clients for specific projects, and internally developed information. Much of the basic gasoline blending data was derived from an update of "U.S. Motor Gasoline Economics, Volume 1, Manufacture of Unleaded Gasoline" published June 1, 1967 by the American Petroleum Institute. In general, the simulation data presently represents a consensus of individual data elements from different sources which have been blended and in some cases modified based on our technical/economic experience in this field.

A. CRUDE OIL PROPERTIES

Table H-1 provides distillation yield data and selected product inspections for the various crude oils (and natural gasoline) used in this study. For most of the crude oils, more than one crude assay was available and the yields presented reflect a composite of the information available. As can be seen in the table, the crude oil assay is divided into individual light hydrocarbons through butanes (methane and ethane are combined together and presented as fuel oil equivalent barrels (FOE)*). Straight-run naphtha is divided into four boiling fractions the lightest of which (C5-160°F) can be isomerized or routed directly to motor gasoline blending.

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*An FOE barrel is equivalent to 6.3×10^6 BTU gross heating value.

The light naphtha fraction (160-200°F) can be reformed or blended to gasoline. The medium naphtha fraction (200-340°F) must be reformed. The heavy naphtha fraction (340-375°F) can be either reformed or blended to middle distillate products.

Straight-run atmospheric gas oil is divided into two fractions - a 375-500°F product with suitable volatility characteristics for blending to aviation turbine fuel and a 500-650°F heating/diesel oil component. Atmospheric bottoms (650°+) is usually processed in a vacuum distillation unit operating at an equivalent cut point of approximately 1,050°F.

Key properties for each crude oil fraction specifically used in the model runs are shown in Tables H-13 and H-14. Other properties of these fractions are also important in product blending, which were met by allowing only suitable boiling range fractions to be blended into these products.

B. PROCESS DATA

Tables H-2 and H-3 provide basic yield data for the catalytic reforming simulations in the model. Product inspections are reported in Table H-15. As noted previously, three different straight-run boiling fractions can be charged to reforming. Unit yields are provided for three severity levels of operation - 90, 95, and 100 clear research octane number (RON) on the C5 + reformate product.

The yields simulate bi-metallic catalyst operation at moderate (300-400 psig) reactor pressure. The hydrogen yield shown in the table and used in the model is the effective hydrogen production available for hydrotreating purposes. This value is approximately 2/3 the stochiometric yield of pure hydrogen and accounts for different operating factors for reforming vs. various hydrotreaters, required purge volume and other inefficiencies in refinery hydrogen usage. The stochiometric hydrogen not made available for hydrotreating purposes is included with the ethane/methane stream to refinery fuel.

Table H-2 presents reforming yield data for straight-run naphthas while Table H-3 presents yield data for catalytic reforming of naphthas supplied from various conversion processes. Two heavy hydrocracked naphthas are identified: one derived from straight-run gas oils (atmospheric or vacuum) feeds and a second produced from cracked gas oil feedstocks (i.e. catalytic cracking, coking, etc.) which produces a higher ring content reformer feedstock. Other potential feeds to catalytic reforming are medium coker naphtha and heavy catalytic cracked gasoline both of which are hydrotreated.

Item Nos. 4, 10, 38, 40, and 43 in the Technical Documentation Reference List provides process yield data on catalytic reforming. The ADL simulation model data is not derived from any specific reference source and the referenced articles are only noted to indicate the range of industry information available and that the selected ADL model data is representative of published plant experience.

The yield data of Table H-2 are a simplified representation of reformer operation, but do provide for ease of computer simulation. Because the reformer simulation is critical to the present study, extensive computer check runs were made with alternate yield data of Tables H-19 and H-20, which represent improved reformer simulation. Check runs were conducted on the Texas Gulf cluster model, which represents a major contribution to the U.S. gasoline production. From these studies, it was concluded that the yields of Table H-2 and H-3 were quite adequate in the present simulation studies.

Table H-4 provides yield data for catalytic cracking. Four different yield structures are displayed for catalytic cracking - at low and high severity operation, and for raw and hydrotreated feed. By utilizing a weighted average of the data, operating severity for untreated feed can range from 65-85% volume conversion, while that for hydrotreated feed can range from 72.5-95%. There are variations in C3/C4 olefin, isobutane, and gasoline yields between cluster models, primarily to balance historical alkylation operating levels. These catalytic cracking yields are based on operation with zeolite catalysts. Item Nos. 5, 14, 15, 17, 38, 42, 43, 44, 46, and 47 in the Technical Documentation Reference List provide process yield data for catalytic cracking.

Hydrocracking yields are presented in Table H-5 for two severity levels of operation - one producing maximum gasoline, and a low severity operation producing approximately 60% jet fuel product. Yields from three different feedstocks are shown for hydrocracking: atmospheric gas oil, vacuum gas oil, and cracked gas oils (from catalytic cracking, coking, etc.). Item Nos. 1, 3, 38, 43, and 45 in the Technical Documentation Reference List provide process yield data for hydrocracking.

Table H-6 provides delayed coking yields for processing vacuum bottoms feedstocks or heavy cycle oil from catalytic cracking. Variations exist in gas oil and coke yields between cluster models to reflect the differing crude slates for these models. Item Nos. 34, 36, 39, and 43 in the Technical Documentation Reference List provide process yield data for delayed coking.

Table H-8 provides yield data for hydrotreating operations while selected product properties are shown in Table H-12. Vacuum and atmospheric bottoms hydrotreating is allowed only in the two grassroots models as it is anticipated that this process will not be commonly installed in existing plants. (See Appendix F) Therefore, data is provided for only two crudes - Arabian Light for the East Coast and Alaskan North Slope for the West Coast. Hydrotreating of Arabian Light atmospheric bottoms is shown producing 1% and .5% sulfur product, and for vacuum bottoms 1% and .6% sulfur. Alaskan North Slope atmospheric bottoms may be desulfurized to .5% and vacuum bottoms to .6%.

Table H-9 provides data on miscellaneous refinery process units, including isomerization, alkylation, and aromatics extraction. Two operating modes are available for isomerization, once-through and recycle. The alkylation unit is assumed to charge a mixed C_3/C_4 olefin stream containing about 1/3 C_3 olefins and 2/3 C_4 olefins. Aromatics extractions charging a reformate produced from light naphtha (100-200°F) feedstock produces a 50/50 mix of BTX/raffinate. The product mix declines to 25/75 when a heavier reformer feedstock is used. Item Nos. 2, 6, 11, 13, 19, 38, and 43 in the Technical Documentation Reference List provide process yield data for the above miscellaneous refinery process units.

Tables H-10 and H-11 provide hydrogen consumption data for hydrotreating and hydrocracking. There are two hydrogen purity systems in the model. A "normal purity" system of approximately 80% hydrogen purity is supplied by catalytic reforming and is suitable for all hydrotreating use. (Only about 2/3 of the stochiometric hydrogen produced by reforming is available for hydrotreating use as noted in the reforming discussion in this section.) Purification of this system for upgrading is not allowed in the model. A "high purity" hydrogen system is required for hydrocracking use which must be supplied by a synthetic hydrogen plant charging either natural gas or naphtha feedstock. Hydrogen consumption for straight-run distillate and residual streams is crude specific, varying with the properties of the respective crude fractions. Item Nos. 28, 29, 38, and 43 in the Technical Documentation Reference List provide process yield data for hydrogen consumption.

C. GASOLINE BLENDING QUALITIES

Tables H-13 through H-16 provide gasoline blending properties for Reid Vapor Pressure (RVP), octanes, and sulfur contents. The original source of much of the RVP and octane data was from the referenced API study noted in the first paragraph in this section. In general, ADL took the respective octane numbers for each non-straight-run component quoted in the study and adjusted for blending bonuses by combining those for premium and regular gasoline in a 1/3-2/3 ratio. The octane numbers thus derived were submitted to an API/NPRA task force as well as to other industry sources for review. Many verbal and written comments were received and the numbers used in this study reflect the consensus of these comments. Item Nos. 7, 8, 9, 16, and 41 in the Technical Documentation Reference List provide industry comments on the gasoline blending octane numbers.

D. SULFUR DISTRIBUTION

The distribution of sulfur in the product streams for several units is presented in Tables H-17 and H-18. Item Nos. 12, 18, 37, and 39 in the Technical Documentation Reference List provide process yield data for sulfur distribution.

E. OPERATING COSTS

Tables H-21 through H-26 provide detailed unit consumptions and cost coefficients of the individual elements that comprise refinery operating costs. These include maintanence, labor, purchased catalysts and chemicals, (including royalties) purchased electricity, steam, cooling water, and refinery fuel. The unit consumptions are provided per barrel of intake for each refinery process, except alkylation, which is per barrel of alkylate.

Refinery steam requirements are balanced by a steam generating facility which consumes additional refinery fuel. Cooling water requirements can be purchased or supplied by a central refinery cooling tower. Unit consumptions for crude and products handling are provided per barrel of total crude oil charged to the refinery, and reflect operating costs associated with receiving and storing crude oil, and blending, storage and loading of finished products.

F. CAPITAL INVESTMENTS

Estimating investments for onsite and offsite process units currently is difficult because of the recent rapid rate of inflation and the long time it takes to build a large, complete petroleum refinery. Nevertheless, investment estimates were necessary and have been determined using data from literature sources, from engineering contractors, and from internal cost files.

In order to minimize the impact of estimates of inflationary factors over the next decade, all investments used in the model were 1975 costs, i.e. all materials and labor were costed on a first quarter, 1975 basis. This is equivalent to a hypothetical case in which a refinery is designed, equipment ordered, and constructed all within the year 1975. In Volume I of the report, these costs are also reported for an assumed level of cost escalation to more nearly reflect the actual costs that will be incurred for the scenarios under evaluation, assumed to be 20%, 17%, 15%, 10%, 10%, 10%, 9%, 9%, 8%, 8%, 8% for the years 1975-1985.

Onsite capital investments were estimated by the unit costs of Table H-27. Variation of the unit investments with unit capacity utilized the standard curves such as reported by Nelson. Item Nos. 20, 21, 22, 27, 30, 31, 32, 33, and 35 in the Technical Documentation Reference List provide typical capital investment numbers.

In assessing refinery costs, units were costed individually using the data of Table H-27. In addition, costs of capacity utilization and severity upgrading were assessed, as discussed in Appendix E. Offsite and working capital requirements for the cluster models were taken to be a constant 40% of onsite costs. Offsite costs and working capital requirements for the grassroots models were obtained by the Nelson complexity factor approach, discussed in detail in Item Nos. 23, 24, 25, and 26 in the Technical Documentation Reference List. With this approach, working capital was taken to be 70% of total onsite capital investment. In the cases considered, offsites and associated costs (including working capital) varied from 200-300% of onsite costs.

In assessing the economic penalties associated with a regulation, a capital-related penalty was taken to be 25%/year of the total capital investment. This capital charge, used for both cluster and grassroots models, provides for return on invested capital, and is equivalent to roughly 12% rate of return on an after tax, discounted cash flow basis. A discussion of capital costs for small refiners appears in Appendix G.

Table H-1. CRUDE AND NATURAL GASOLINE YIELDS; CRUDE PROPERTIES

	Domestic crudes						Foreign crudes						
Yield, Volume %	Louisiana	West Texas Sour	Oklahoma	California Wilmington	California Ventura	Alaskan North Slope	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana	Algerian Hassi Messaoud	Mixed Canadian	Indonesian Minas	Natural gasoline
Methane/ethane (FOE)	–	.04	–	.002	.005	.06	.04	–	.01	.04	.05	.003	–
Propane	.20	.50	.48	.093	.400	.28	.04	.17	.59	1.21	1.13	.139	–
Isobutane	.40	.40	.46	.096	.295	.13	.51	.17	.27	.53	.49	.141	4.71
Normal butane	.70	1.30	2.18	.318	1.069	.49	.79	1.06	.45	3.27	1.98	.379	7.58
Straight-run naphtha:													
C5-160°F	4.09	5.70	7.70	2.09	5.72	3.53	2.70	4.86	3.89	8.29	6.60	1.70	62.02
Light 160-200°F	2.99	3.80	4.76	2.47	3.25	2.57	3.40	3.27	2.20	5.00	3.89	1.80	13.99
Medium 200-340°F	13.05	15.50	16.66	7.64	14.26	9.25	11.70	14.83	9.09	21.19	16.50	9.10	11.70
Heavy 340-375°F	3.47	3.30	3.83	1.50	3.34	2.09	2.80	3.86	2.62	5.02	3.51	2.60	–
Light gas oil 375-500°F	17.50	13.39	13.76	9.29	11.54	12.43	18.10	13.39	10.30	15.78	14.40	10.70	–
Heavy gas oil 500-650°F	19.50	14.11	11.77	11.96	12.53	15.50	20.80	15.01	12.70	11.92	15.20	16.00	–
Vacuum overhead 650-1050°F	32.50	29.60	28.04	38.54	32.06	29.46	30.40	29.50	32.80	22.71	25.60	41.00	–
Vacuum bottoms 1050°F+	5.60	12.30	10.25	26.00	16.50	23.52	8.50	13.70	25.00	4.99	10.60	18.00	–
Crude properties													
Gravity (°API)	36.2	33.4	40.2	19.6	29.7	27.5	29.4	34.5	26.3	44.7	39.0	35.3	–
% Weight sulfur	.22	1.63	.212	1.28	1.86	.96	.21	1.7	1.51	.13	.56	.07	–

Table H-2. YIELD DATA

Reforming of SR Naphtha

90 RON Severity

Stream[a]	Domestic crudes						Foreign crudes						
	Louisiana	West Texas Sour	Oklahoma	Calif. Wilmington	Calif. Ventura	Alaskan North Slope	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana	Algerian Hassi Massoud	Mixed Canadian	Indonesian Minas	Natural Gasoline
Light feed (160-200°)													
Hydrogen (MSCF)[b]	.546	.660	.680	.748	.680	.442	.719	.349	.680	.546	.704	.442	.546
Ethane/methane (FOE)	.0526	.0496	.0496	.0515	.0496	.0564	.0495	.0802	.0496	.0626	.0495	.0564	.0526
Propane	.0437	.0291	.0267	.0069	.0267	.0622	.0220	.0819	.0291	.0437	.0238	.0622	.0437
Isobutane	.0192	.0111	.0097	.001	.0097	.0296	.007	.0405	.0111	.0192	.008	.0296	.0192
Normal butane	.0336	.0218	.0199	.0061	.0199	.0486	.016	.0640	.0218	.0336	.0175	.0486	.0336
90 Reformate	.8449	.8808	.8888	.9284	.8886	.7993	.9049	.7552	.8808	.8449	.8069	.7993	.8449
Medium feed (200-340°)													
Hydrogen (MSCF)[b]	.660	.748	.660	.852	.790	.790	.852	.442	.704	.660	.790	.442	.660
Ethane/methane (FOE)	.0496	.0515	.0496	.0496	.0507	.0507	.0496	.0564	.0506	.0496	.0507	.0564	.0496
Propane	.0291	.0069	.0291	.0065	.0067	.0067	.0065	.0669	.018	.0291	.0067	.0622	.0291
Isobutane	.0111	.001	.0111	—	.0006	.0006	—	.0318	.006	.0111	.0006	.0296	.0111
Normal butane	.0218	.0061	.0218	.0033	.006	.006	.0033	.0487	.014	.0218	.005	.0486	.0218
90 Reformate	.8808	.9284	.8808	.9576	.9401	.9401	.9576	.7993	.9046	.8808	.9401	.7993	.8808
Heavy feed (340-375°)													
Hydrogen (MSCF)[b]	.660	.748	.660	.852	.790	.790	.852	.442	.704	.660	.790	.442	
Ethane/methane (FOE)	.0496	.0515	.0496	.0496	.0507	.0507	.0496	.0494	.0506	.0496	.0507	.0564	
Propane	.0291	.0069	.0291	.0065	.0067	.0067	.0065	.0669	.018	.0291	.0067	.0622	
Isobutane	.0111	.001	.0111	.0001	.0006	.0006	.0001	.0318	.006	.0111	.0006	.0296	
Normal butane	.0218	.0061	.0218	.0032	.005	.005	.0032	.0487	.014	.0218	.005	.0486	
90 Reformate	.8808	.9284	.8808	.9576	.9401	.9401	.9576	.7993	.9046	.8808	.9401	.7993	

[a] LV fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

Table H-2 (continued). YIELD DATA

Reforming of SR Naphtha

95 RON Severity

Stream[a]	Domestic crudes						Foreign crudes						
	Louisiana	West Texas Sour	Oklahoma	Calif. Wilmington	Calif. Ventura	Alaskan North Slope	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana	Algerian Hassi Messaoud	Mixed Canadian	Indonesian Minas	Natural Gasoline
Light feed (160-200°)													
Hydrogen (MSCF)[b]	.546	.660	.680	.748	.680	.442	.719	.349	.660	.546	.704	.442	.546
Ethane/methane (FOE)	.066	.0656	.0656	.0675	.0656	.0654	.0655	.0646	.0656	.066	.0655	.0654	.066
Propane	.064	.0441	.0405	.0219	.0405	.0819	.0333	.0983	.0441	.064	.036	.0819	.064
Isobutane	.027	.0161	.0141	.004	.0141	.0368	.0102	.0457	.0161	.027	.0117	.0368	.027
Normal butane	.043	.0288	.0263	.0131	.0263	.0557	.0212	.0673	.0288	.043	.0231	.0557	.043
95 Reformate	.80	.8428	.8505	.8904	.8505	.7613	.8659	.7262	.8428	.80	.8601	.7613	.80
Medium feed (200-340°)													
Hydrogen (MSCF)[b]	.660	.748	.660	.852	.790	.790	.852	.442	.704	.660	.790	.442	.660
Ethane/methane (FOE)	.0656	.0675	.0656	.0656	.0667	.0667	.0656	.0654	.0666	.0656	.0667	.0654	.0656
Propane	.0441	.0219	.0441	.0099	.0171	.0171	.0099	.0819	.033	.0441	.0171	.0819	.0441
Isobutane	.0161	.004	.0161	–	.0024	.0024	–	.0368	.01	.0161	.0024	.0368	.0161
Normal butane	.0288	.0131	.0288	.0044	.0096	.0096	.0044	.0557	.021	.0288	.0096	.0557	.0288
95 Reformate	.8428	.8904	.8428	.9163	.9008	.9006	.9163	.7613	.8666	.8428	.9008	.7613	.8428
Heavy feed (340-375°)													
Hydrogen (MSCF)[b]	.660	.748	.660	.852	.790	.790	.852	.442	.704	.660	.790	.442	
Ethane/methane (FOE)	.0656	.0675	.0656	.0656	.0667	.0667	.0656	.0654	.0666	.0656	.0667	.0654	
Propane	.0441	.0219	.0441	.0099	.0171	.0171	.0099	.0819	.033	.0441	.0171	.0819	
Isobutane	.0161	.004	.0161	–	.0024	.0024	–	.0368	.01	.0161	.0024	.0368	
Normal butane	.0288	.0131	.0288	.0044	.0096	.0096	.0044	.0557	.021	.0288	.0096	.0557	
95 Reformate	.8428	.8904	.8428	.9163	.9008	.9008	.9163	.7613	.8666	.8428	.9008	.7613	

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[a] LV fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

Table H-2 (continued). YIELD DATA

Reforming of SR Naphtha

100 RON Severity

Stream[a]	Domestic crudes						Foreign crudes							
	Louisiana	West Texas Sour	Oklahoma	Calif. Wilmington	Calif. Ventura	Alaskan North Slope	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana	Algerian Hassi Messaoud	Mixed Canadian	Indonesian Minas	Natural Gasoline
Light feed (100-200°)													
Hydrogen (MSCF)[b]	.596	.754	.750	.838	.750	.487	.772	.394	.754	.596	.767	.487	.596
Ethane/methane (FOE)	.080	.0814	.0813	.0813	.0813	.0788	.0809	.0784	.0814	.080	.0811	.0788	.080
Propane	.090	.0715	.0681	.0508	.0681	.1066	.0614	.1218	.0715	.090	.0639	.1066	.090
Isobutane	.035	.0245	.0226	.0129	.0226	.0444	.0189	.0530	.0245	.035	.0203	.0444	.035
Normal butane	.054	.0404	.038	.0253	.0380	.0662	.0331	0773	.0404	.054	.0349	.0662	.054
100 Reformate	.745	.7848	.792	.8292	.7920	.7089	.8064	.6762	.7848	.745	.8010	.7089	.745
Medium feed (200-340°)													
Hydrogen (MSCF)[b]	.7520	.838	.7520	.910	.867	.867	.910	.487	.795	.7520	.867	.487	.7520
Ethane/methane (FOE)	.0815	.0813	.0815	.082	.0816	.0816	082	.0788	.0814	.0815	.0816	.0788	.0815
Propane	.0715	.0508	.0715	.0397	.0464	.0464	.0397	1066	.0612	.0715	.0464	1066	.0715
Isobutane	.0245	0129	.0245	.0065	.0103	.0103	.0065	.0444	.0187	.0245	.0103	.0444	.0245
Normal butane	.0404	.0253	.0404	.017	.0220	.022	.017	.0662	.0329	.0404	.0220	.0662	.0404
100 Reformate	.7848	.8292	.7848	.8533	.8388	.8386	.8533	.7089	.807	.7848	.8388	.7089	.7848
Heavy feed (340-375°)													
Hydrogen (MSCF)[b]	.7520	.838	.7520	.910	.867	.867	.910	.487	.795	.7520	.867	.487	
Ethane/methane (FOE)	.0815	.0813	.0815	.082	.0816	.0816	.082	.0788	.0814	.0815	.0816	.0788	
Propane	.0715	.0508	.0715	.0397	.0464	.0464	.0397	.1066	.0612	.0715	.0464	.1066	
Isobutane	.0245	.0129	.0245	.0065	.0103	.0103	.0065	.0444	.0187	.0245	.0103	.0444	
Normal butane	.0404	.0253	.0404	.017	.0220	.022	.017	.0862	.0329	.0404	.0220	.0662	
100 Reformate	.7848	.8292	.7848	.8533	.8388	.8389	.8533	.7089	.807	.7848	.8388	.7089	

[a] LV fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

Table H-3. YIELD DATA

Reforming of Conversion Naphtha

Stream[a]	Heavy hydrocracked naphtha		Medium coker naphtha	Heavy cat. naphtha
	Straight run gas oil feed[c]	Cracked gas oil feed		
90 RON Severity				
Hydrogen (MSCF)[b]	.720	.852	.660	.660
Ethane/methane (FOE)	.0495	.0496	.0496	.0496
Propane	.0220	.0065	.0291	.0291
Isobutane	.007	–	.0111	.0111
Normal butane	.016	.0033	.0218	.0218
90 Reformate	.9049	.9573	.8808	.8808
95 RON Severity				
Hydrogen (MSCF)[b]	.720	.852	.660	.660
Ethane/methane (FOE)	.0655	.0656	.0656	.0656
Propane	.0333	.0099	.0441	.0441
Isobutane	.0102	–	.0161	.0161
Normal butane	.0212	.0044	.0288	.0288
95 Reformate	.8659	.9163	.8428	.8428
100 RON Severity				
Hydrogen (MSCF)[b]	.774	.910	.753	.753
Ethane/methane (FOE)	.0808	.082	.0814	.0814
Propane	.0614	.0397	.0715	.0715
Isobutane	.0189	.0065	.0245	.0245
Normal butane	.0331	.017	.0404	.0404
100 Reformate	.8064	.8533	.7848	.7848

[a] LV fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

[c] Includes atmospheric gas oils and vacuum overhead.

Table H-4. YIELD DATA

Catalytic Cracking

(LV fraction on feed)

Stream	Low sev cat cracking								High sev cat cracking							
	Louisiana	Texas	Large Midw.	Small Midw.	East Coast	West Coast	East Grassroots	West Grassroots	Louisiana	Texas	Large Midw.	Small Midw.	East Coast	West Coast	East Grassroots	West Grassroots
Untreated 850-1050°F feed																
Methane/ethane (FOE)	.075	.025	.025	.025	.025	.025	.025	.025	.048	.048	.048	.048	.048	.048	.048	.048
C_3/C_4 Olefins	123	.128	.088	.128	.088	.088	.128	.128	.178	.178	.158	.178	178	178	.178	178
Propane	018	.013	.016	.013	.013	.013	.013	.013	.03	.03	.033	.03	.03	.03	.03	.03
Isobutane	046	.046	.083	.066	.061	.051	.046	.046	.09	.09	.103	.11	.09	.09	.09	.09
Normal butane	008	.008	.008	.008	.008	.008	.008	.008	.022	.022	.022	.022	.022	.022	.022	.022
Cat. gasoline (C_5 to 430°F)	.52	.52	.52	.50	.538	.538	.52	.52	.60	.60	.60	.58	.60	.60	.60	.60
Light cycle oil	27	27	.27	.27	.27	.27	.27	.27	.10	.10	.10	.17	.10	10	.10	10
Heavy cycle oil	08	08	.08	.11	.08	.08	.08	.08	.05	.05	.05	.05	.05	.05	.05	.05
Treated 850-1050°F feed																
Methane/ethane (FOE)	018	.018	.018	.018	.018	.018	.018	018	[a]	[a]	[a]	[a]	[a]	[a]	.033	.031
C_3/C_4 Olefins	147	.137	.137	.137	.137	.137	.137	137							.191	191
Propane	020	020	.020	.020	020	.020	.020	.020							.043	043
Isobutane	07	.07	.07	.08	.07	.07	.07	07							.128	128
Normal butane	012	.012	.012	.012	.012	.012	.012	.012							.031	.031
Cat. gasoline (C_5 to 430°F)	57	58	58	57	58	58	58	58							089	089
Light cycle oil	212	212	212	212	.212	.212	.212	212							033	033
Heavy cycle oil	063	063	063	063	063	063	.063	063							017	017

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[a] High severity catalytic cracking of hydrotreated feed is not used in the cluster models.

Table H-5. YIELD DATA

Hydrocracking

(LV fraction on feed)

Stream	High severity			Medium severity		
	Heavy G.O.	Vacuum G.O.	Cracked G.O.	Heavy G.O.	Vacuum G.O.	Cracked G.O.
Ethane/methane (FOE)	.004	.0057	.005	.003	.004	.003
Propane	.054	.0768	.067	.040	.050	.045
Isobutane	.144	.1520	.117	.080	.084	.070
Normal butane	.060	.0630	.052	.045	.048	.040
Light gasoline	.364	.382	.317	.220	.231	.200
Heavy naphtha	.680	.692	.780	.252	.265	.312
Jet fuel				.600	.630	.610

Table H-6. YIELD DATA
Coking
(LV fraction on feed)

Stream	Vacuum bottoms feed						Heavy cycle oil feed					
	Louisiana	Texas	Large Midw.	Small Midc.	West Coast	West Grassroots	Louisiana	Texas	Large Midw.	Small Midc.	West Coast	West Grassroots
Methane/ethane (FOE)	.095	.095	.095	.095	.095	.095	.0743	.0743	.0743	.0743	.0743	.0743
Propane	.02	.001	.02	.001	.001	.001	.0007	.0007	.015	.0007	.0007	.0007
C_3/C_4 Olefins	.039	.039	.039	.039	.039	.039	.030	.030	.030	.030	.030	.030
Isobutane	.008	.008	.008	.008	.008	.008	.006	.006	.006	.006	.006	.006
Normal butane	.022	.022	.022	.022	.022	.022	.017	.017	.017	.017	.017	.017
Light coker naphtha	.105	.105	.105	.105	.105	.105	.080	.080	.080	.080	.080	.080
Med. coker naphtha	.187	.187	.187	.187	.187	.187	.142	.142	.142	.142	.142	.142
Coker gas oil	.410	.413	.381	.308	.428	.428	.594	.5932	.560	.560	.560	.560
Coke	.261	.258	.290	.363	.243	.243	.1275	.1283	.1615	.1615	.1615	.1615

Table H-7. YIELD DATA
Visbreaking
(LV fraction on feed)

Stream	West Coast cluster[a]
1050°F+ feed	
Methane/ethane (FOE)	.0168
Propane	.0002
Isobutane	.002
Normal butane	.005
C_3/C_4 Olefins	.005
Visbreaker naphtha	.110
Visbreaker gas oil	.4020
Tar	.4900

[a] Visbreaking not used in any other cluster or grassroots model.

Table H-8. YIELD DATA

Desulfurization
(LV Fraction on feed)

Stream	SR naph	SR gas oil	Coker naph.	Vacuum OVHD	Light cycle oil	Cat naph	Atm. bottoms to 0.5% North Slope	Atm. bottoms to 1.0% Arab Light	Atm. bottoms to 0.5% Arab Light	Vac. bottoms to 0.6% North Slope	Vac. bottoms to 1.0% Arab Light	Vac. bottoms to 0.8% Arab Light
Ethane/methane (FOE)				.001			.0020	.0030	.0031	.0041	.0041	.0049
Propane		.001		.001	.001		.0024	.0030	.0046	.0041	.0041	.0072
Isobutane		.001		.002	.001		.0016	.0020	.0025	.0028	.0028	.0039
Normal butane		.001		.002	.001		.0016	.0020	.0025	.0028	.0028	.0039
Naphtha		.008		.01	.008		.0089	.0070	.0150	.0160	.0160	.0145
Desulfurized product	1.0	.990	1.01	.995	.990		1.0051	.9934	.9966	1.008	1.008	1.022
Light gasoline						.672						
Heavy gasoline						.336						

Table H-9. YIELD DATA

Miscellaneous Process Units

Stream[a]	Isomerization		Alkylation	Aromatics extraction		Hydrogen from natural gas (FOE) feed	Sulfur recovery	
	Single	Recycle	C_3/C_4 Olefins[b]	160-200°F feed	200-340°F feed		95% (Wgt fraction on feed)	99.95% (Wgt fraction on feed)
Ethane/methane (FOE)	0.023	0.036						
Isomerate	0.965	0.945						
Alkylate			1.77					
BTX aromatics				0.5	0.25			
Raffinate				0.5	0.75			
Hydrogen (MSCF/Bbls)						23.9		
Sulfur							0.894	0.941
Sulfur oxides							0.094	0.009

[a] LV fraction on feed unless otherwise noted.

[b] Isobutane consumption — 1.19 per unit volume C_3/C_4 olefin feed.

Table H-10. HYDROGEN CONSUMPTION DATA

Desulfurization of Crude-specific Streams

(MSCF/Bbl feed)

Stream	Domestic crudes						Foreign crudes					
	Louisiana	West Texas Sour	Oklahoma	California Wilmington	California Ventura	Alaskan North Slope	Mixed Canadian	Arabian Light	Nigerian Forcados	Algerian Hassi Messaoud	Venezuelan Tia Juana	Indonesian Minas
Normal purity hydrogen												
SR naphtha	.105	.105	.105	.105	.105	.105	.105	.105	.105	.105	.105	.105
Light gas oil	.140	.190	.140	.140	.140	.140	.170	.170	—a—	.150	.170	.140
Heavy gas oil	.150	.250	.150	.150	.150	.150	.250	.250	—a—	—a—	.250	.150
Vacuum overhead	.300	.300	.300	.300	.300	.300	.300	.300	.300	.300	.300	.300
High purity hydrogen												
Atmospheric bottoms												
to 1.0% sulfur								.530				
to .5% sulfur						.330		.600				
Vacuum bottoms												
to 1.0% sulfur								.860				
to .6% sulfur						.860		1.060				

[a] Not desulfurized.

Table H-11. HYDROGEN CONSUMPTION DATA

Hydrocracking and Desulfurization of Model-Specific Streams

(MSCF/Bbl feed)

Stream	Louisiana	Texas	Small Midc.	Large Midwest	West Coast	East Coast	West Grassroots	East Grassroots -sour	East Grassroots -sweet
Normal purity hydrogen									
Light cycle oil	.220	.220	.220	.220	.220	.220	.220	.220	.220
Coker naphthas	.600	.600	.600	.600	.600	.600	.600	.600	.600
Cat. gasoline	.600	.600	.600	.600	.600	.600	.600	.600	.600
High purity hydrogen									
High sev hydrocracking									
Heavy gas oil feed	1.90	1.95	1.90	1.95	1.95	1.95	1.95	1.95	1.90
Vacuum gas oil feed	2.45	2.50	2.45	2.50	2.50	2.50	2.50	2.50	2.45
Cracked gas oil feed	3.10	3.15	3.10	3.15	3.15	3.15	3.15	3.15	3.10
Medium sev hydrocracking									
Heavy gas oil feed	1.70	1.75	1.70	1.75	1.75	1.75	1.75	1.75	1.70
Vacuum gas oil feed	1.83	1.88	1.83	1.88	1.88	1.88	1.88	1.88	1.83
Cracked oil gas feed	2.70	2.75	2.70	2.75	2.75	2.75	2.75	2.75	2.70

Table H-12. SULFUR REMOVAL

Levels in Desulfurization

Stream	Effluent Stream sulfur, %[a]	Comment
Isomerization feed		Level required for isomerization feed.
C_5 to 160°F	1 PPM	
Reformer feed		Level required for reformer feed.
Light SR naphtha (160-200°F)	1 PPM	
Medium SR naphtha (200-340°F)	1 PPM	
Heavy SR naphtha (340-375°F)	1 PPM	
Heavy hydrocrackate	1 PPM	
Heavy cat. naphtha	1 PPM	
Coker naphtha	1 PPM	
Cat. feed		
Vacuum overhead	0.2	Desulfurization to 0.2% wt.S or 85% feed sulfur removal (whichever is lower).
Heavy atomospheric gas oil	0.2	
Coker gas oil	0.2	
Other streams		
Light atmospheric gas oil	1% feed sulfur	
Heavy atmospheric gas oil	5% feed sulfur	
Cat. cycle oil	15% feed sulfur	
Atmospheric bottoms		
– Alaskan North Slope	0.5	
– Arabian Light	1.0/0.5	Two processes (corresponding to the two sulfur levels listed) exist in model.
Vacuum bottoms		
– Alaskan North Slope	0.6	
– Arabian Light	1.0/0.6	Two processes (corresponding to the two sulfur levels listed) exist in model.

[a] Percent, unless otherise noted.

Table H-13. STREAM QUALITIES

Domestic Crudes

Stream	Specific gravity	Sulfur content, % weight	Viscosity[a]	Smoke point, mm.	Gasoline blending qualities							Mid-fill blend number, % distilled at	
						RON			MON				
					R.V.P.	Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	150°F	210°F
Louisiana													
Full range naphtha	.7511	.0072											
C5 to 200°F	.693	.0021			8.1	70.8	78.2	88.8	68.5	76.8	88.2	41.0	95.0
C5 to 160°F	.668	.0002			11.0	73.0	80.6	91.0	71.2	79.6	90.8	91.3	100.0
Light 160-200°F	.727	.0045			4.1	67.8	75.3	85.8	64.8	73.0	84.6	0.0	99.0
Medium 200-340°F	.762	.0093											
Heavy 340-375°F	.816	.0089	4.53	22.5									
Atmospheric gas oil 375-650°F	.837	.0649	15.18										
Light gas oil 375-500°F	.8220	.0362	9.04	20.0									
Heavy gas oil 500-650°F	.8504	.0901	20.68										
Atmospheric bottoms 650°F+	.9108	.4175	31.74										
Vacuum overhead 650-1050°F	.8974	.3221	28.75										
Vacuum bottoms 1050°F+	.9881	.9207	49.14										
Once through isomerate	.65	.0001			11.5	82.1	88.0	96.0	82.8	87.8	94.9	93.0	105.0
Recycle isomerate	.64	.0001			12.0	90.0	94.0	99.9	86.8	92.5	100.2	95.0	100.0
West Texas Sour													
Full range naphtha	.7587	.1496											
C5 to 200°F	.691	.0364			8.9	67.8	77.0	89.3	58.6	67.2	84.8	51.0	98.0
C5 to 160°F	.664	.0288			11.5	72.1	79.8	90.5	65.7	74.3	86.2	91.3	100.0
Light 160-200°F	.732	.0466			5.0	61.4	72.5	87.5	48.0	59.0	82.7	0.0	99.0
Medium 200-340°F	.793	.1610											
Heavy 340-375°F	.794	.3790	4.53	22.0									
Atmospheric gas oil 375-650°F	.8440	.9146	14.44										
Light gas oil 375-500°F	.8251	.5787	8.32	18.0									
Heavy gas oil 500-650°F	.8633	1.2187	20.26										
Atmospheric bottoms 650°F+	.9467	2.2145	35.09										
Vacuum overhead 650-1050°F	.9167	1.8513	28.51										
Vacuum bottoms 1050°F+	1.0187	3.0018	50.92										
Once through isomerate	.65	.0001			12.0	81.0	86.9	95.0	79.0	83.9	91.0	93.0	100.0
Recycle isomerate	.64	.0001			12.5	89.0	93.0	98.9	85.0	90.5	98.0	95.0	100.0

[a] Refutas blending values for viscosity in centistokes @122°F.

Table H-13 (continued). STREAM QUALITIES

Domestic Crudes

Stream	Specific gravity	Sulfur content, % weight	Viscosity[a]	Smoke point, mm.	Gasoline blending qualities							Mid-fill blend number, % distilled at	
						RON			MON				
					R.V.P.	Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	150°F	210°F
Oklahoma													
Full range naphtha	.7315	.0074											
C5 to 200°F	.665	.0022			10.2	70.3	77.5	87.7	69.6	76.3	85.9	41.0	95.0
C5 to 160°F	.643	.0002			14.2	70.7	78.4	88.9	69.7	77.0	87.0	91.3	100.0
Light 160-200°F	.701	.0051			3.7	69.7	76.4	85.8	69.4	75.4	84.1	0.0	99.0
Medium 200-340°F	.763	.0105											
Heavy 340-375°F	.809	.0090	4.53	–									
Atmospheric gas oil 375-650°F	.838	.0912	15.18										
Light gas oil 375-500°F	.825	.0572	9.04	–									
Heavy gas oil 500-650°F	.854	.1296	20.68										
Atmospheric bottoms 650°F+	.9080	.3306	31.74										
Vacuum overhead 650-1050°F	.8970	.2327	28.75										
Vacuum bottoms 1050°F+	.9380	.5883	49.14										
Once through isomerate	.65	.0001			14.7	80.3	86.0	93.7	81.8	85.8	91.8	93.0	105.0
Recycle isomerate	.64	.0001			15.2	88.7	92.5	98.1	86.4	91.5	98.4	95.0	100.0
California Wilmington													
Full range naphtha	.763	.0472											
C5 to 200°F	.689	.0098			6.4	83.1	87.8	94.6	79.7	84.8	92.1	41.0	95.0
C5 to 160°F	.643	.0100			10.5	85.9	90.2	96.5	82.3	87.4	94.3	91.3	100.0
Light 160-200°F	.728	.0097			2.9	80.7	85.8	93.0	77.5	82.7	90.2	0.0	99.0
Medium 200-340°F	.794	.0446											
Heavy 340-375°F	.832	.1543	3.71	22.9									
Atmospheric gas oil 375-650°F	.881	.6867	19.0										
Light gas oil 375-500°F	.860	.3823	11.78	16.3									
Heavy gas oil 500-650°F	.898	.9124	24.89										
Atmospheric bottoms 650°F+	.9929	1.6557	44.4										
Vacuum overhead 650-1050°F	.966	1.3126	33.84										
Vacuum bottoms 1050°F+	1.033	2.1311	58.34										
Once through isomerate	.637	.0001			11.0	89.0	94.1	101.4	87.5	91.3	96.8	93.0	105.0
Recycle isomerate	.635	.0001			11.5	93.1	97.0	103.3	89.1	94.2	101.6	95.0	100.0

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[a] Refutas blending values for viscosity in centistokes @122°F.

Table H-13 (continued). STREAM QUALITIES

Domestic Crudes

Stream	Specific gravity	Sulfur content, % weight	Viscosity[a]	Smoke point, mm.	Gasoline blending qualities							Mid-fill blend number, % distilled at	
						RON			MON				
					R.V.P.	Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	150°F	210°F
California Ventura													
Full range naphtha	.7471	.1006											
C5 to 200°F	.6824	.0209			8.9	78.8	85.2	94.0	74.8	81.5	91.0	41.0	95.0
C5 to 160°F	.6429	.0200			11.9	80.4	86.5	94.7	76.2	82.7	91.8	91.3	100.0
Light 160-200°F	.7519	.0224			3.6	76.0	83.0	92.8	72.3	79.5	89.6	0.0	99.0
Medium 200-340°F	.7746	.1019											
Heavy 340°-375°F	.8034	.2769	3.25	24.1									
Atmospheric gas oil 375-650°F	.8521	.7334	11.4										
Light gas oil 375-500°F	.8392	.2797	7.09	19.0									
Heavy gas oil 500-650°F	.8640	1.1393	14.9										
Atmospheric bottoms 650°F+	.9676	2.4672	37.39										
Vacuum overhead 650-1050°F	.9350	1.5411	31.9										
Vacuum bottoms 1050°F+	1.035	4.1959	48.76										
Once through isomerate	.6399	.0001			12.4	86.7	92.7	101.0	85.6	89.7	95.5	93.0	105.0
Recycle isomerate	.6379	.0001			12.9	92.3	96.5	102.9	87.8	93.2	100.7	95.0	100.0
Alaskan North Slope													
Full range naphtha	.7518	.0177											
C5 to 200°F	.6801	.0147			7.0	72.2	78.0	86.4	70.3	76.5	85.5	41.0	96.0
C5 to 160°F	.6526	.0105			8.7	75.5	81.0	89.0	73.6	79.6	88.1	91.3	100.0
Light 160-200°F	.7179	.0199			4.7	67.7	74.0	82.8	65.8	72.4	81.9	0.0	99.0
Medium 200-340°F	.7807	.0160											
Heavy 340-375°F	.8151	.0291	1.7	21.85									
Atmospheric gas oil 375-650°F	.8601	.3281	12.5										
Light gas oil 375-500°F	.8349	.1615	8.3	17.9									
Heavy gas oil 500-650°F	.8803	.4547	15.9										
Atmospheric bottoms 650°F+	.9581	1.5310	39.4										
-Desulf to .5% wgt sulfur	.937	.5000	36.5										
Vacuum overhead 650-1050°F	.9281	1.1029	33.2										
Vacuum bottoms 1050°F+	.9957	2.0313	47.3										
-Desulf to .6% wgt sulfur	.946	.6000	43.0										
Once through isomerate	.645	.0001			9.2	83.6	87.8	93.8	84.3	87.7	92.7	93.0	100.0
Recycle isomerate	.64	.0001			9.7	90.6	93.7	98.2	87.3	92.2	98.9	96.0	100.0

[a] Reflects blending values for viscosity in centistokes @122°F.

Table H-14. STREAM QUALITIES

Foreign Crudes and Natural Gasoline

Stream	Specific gravity	Sulfur content, % weight	Viscosity[a]	Smoke point, mm.	Gasoline blending qualities							Mid-fill blend number, % distilled at	
					R.V.P.	RON			MON				
						Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	158°F	212°F
Nigerian Forcados													
Full range naphtha	.762	.0072											
C5 to 200°F	.702	.0001			8.1	77.1	81.8	89.0	73.5	79.0	87.0	41.0	95.0
C5 to 160°F	.670	.0001			11.0	81.0	86.0	92.8	77.0	82.5	90.0	91.3	100.0
Light 160-200°F	.727	.0001			5.6	74.0	78.7	86.0	70.7	76.3	84.6	0.0	99.0
Medium 200-340°F	.780	.0053											
Heavy 340-375°F	.816	.0281	4.53	22.5									
Atmospheric gas oil 375-650°F	.874	.1452	14.14										
Light gas oil 375-500°F	.854	.0790	9.38	21.0									
Heavy gas oil 500-650°F	.891	.2015	18.15										
Atmospheric bottoms 650°F+	.954	.3845	35.00										
Vacuum overhead 650-1050°F	.942	.3125	31.25										
Vacuum bottoms 1050°F+	.998	.6286	47.85										
Once through isomerate	.667	.0001			11.5	87.0	92.5	100.0	86.0	89.3	94.2	93.0	105.0
Recycle isomerate	.665	.0001			12.0	92.5	96.3	102.2	88.0	93.0	99.8	95.0	100.0
Arabian Light													
Full range naphtha	.7335	.0292											
C5 to 200°F	.669	.0231			9.0	60.5	65.0	80.8	59.4	67.0	78.7	53.5	100.0
C5 to 160°F	.657	.0220			11.0	66.0	74.0	85.6	64.5	71.9	82.2	91.3	100.0
Light 160-200°F	.686	.0247			6.0	52.3	58.5	73.7	51.8	57.9	73.5	0.0	99.0
Medium 200-340°F	.7554	.0266											
Heavy 340-375°F	.797	.0487	4.53	27.0									
Atmospheric gas oil 375-650°F	.8278	.6849	11.786										
Light gas oil 375-500°F	.8072	.2203	7.09	23.0									
Heavy gas oil 500-650°F	.8483	1.0807	15.98										
Atmospheric bottoms 650°F+	.9484	3.0820	34.31										
–Desulf. to 1.0% sulfur	.920	1.000	32.0										
–Desulf. to .5% sulfur	.9117	.5000	30.86										
Vacuum overhead 650-1050°F	.9154	2.3215	28.51										
Vacuum bottoms 1050°F+	1.0195	4.5530	46.79										
–Desulf. to 1.0% sulfur	.9567	1.000	46.79										
–Desulf. to .6% sulfur	.9478	.6000	46.79										
Once through isomerate	.654	.0001			11.5	78.0	83.6	91.5	78.2	81.8	87.1	93.0	105.0
Recycle isomerate	.652	.0001			12.0	87.0	93.0	97.0	84.5	89.2	96.0	95.0	100.0

[a] Refutas blending values for viscosity in centistokes @122°F.

Table H-14. (continued). STREAM QUALITIES

Foreign Crudes and Natural Gasoline

Stream	Specific gravity	Sulfur content, % weight	Viscosity[a]	Smoke point, mm.	Gasoline blending qualities							Mid-fill blend number, % distilled at	
					R.V.P.	RON			MON				
						Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	150°F	210°F
Venezuelan Tia Juana													
Full range naphtha	.7362	.0190											
C5 to 200°F	.682	.0078			10.6	67.2	76.0	88.0	67.2	78.4	88.7	76.0	100.0
C5 to 160°F	.659	.0046			14.0	70.3	79.8	90.7	70.3	79.3	91.4	91.3	100.0
Light 160-200°F	.723	.0129			4.6	61.7	69.7	83.2	61.7	71.0	83.9	0.0	98.0
Medium 200-340°F	.7628	.0238											
Heavy 340-375°F	.7708	.0261	3.25	25.0									
Atmospheric gas oil 375-650°F	.8473	.4580	11.4										
Light gas oil 375-500°F	.826	.1900	7.09	21.0									
Heavy gas oil 500-650°F	.865	.6890	14.9										
Atmospheric bottoms 650°F+	.966	2.1999	39.07										
Vacuum overhead 650-1050°F	.922	1.6292	29.73										
Vacuum bottoms 1050°F+	1.0236	2.8743	51.32										
Once through isomerate	.65	.0001			14.5	80.1	86.6	95.3	82.2	87.7	96.2	93.0	106.0
Recycle isomerate	.64	.0001			15.0	88.5	93.0	99.3	86.5	92.5	100.5	95.0	100.0
Algerian Hassi Messaoud													
Full range naphtha	.738	.0070											
C5 to 200°F	.678	.0021			8.2	65.0	73.0	84.2	61.5	70.8	83.6	57.0	100.0
C5 to 160°F	.657	.0002			9.9	68.7	76.5	87.5	66.8	74.9	86.3	91.3	100.0
Light 160-200°F	.716	.0051			5.4	58.9	65.0	78.7	52.7	60.7	78.1	0.0	98.0
Medium 200-340°F	.764	.0091											
Heavy 340-375°F	.788	.0082	1.76	22.0									
Atmospheric gas oil 375-650°F	.834	.0449	9.38										
Light gas oil 375-500°F	.810	.0201	3.25	24.0									
Heavy gas oil 500-650°F	.866	.0756	17.51										
Atmospheric bottoms 650°F+	.910	.3502	30.48										
Vacuum overhead 650-1050°F	.892	.2249	28.97										
Vacuum bottoms 1050°F+	.990	.8865	46.49										
Once through isomerate	.654	.0001			10.4	79.2	84.9	92.7	79.8	84.5	91.1	93.0	105.0
Recycle isomerate	.652	.0001			10.9	87.9	91.9	97.6	85.4	90.7	98.0	95.0	100.0

[a] Refutas blending values for viscosity in centistokes @122°F.

Table H-14. (continued). STREAM QUALITIES

Foreign Crudes and Natural Gasoline

					Gasoline blending qualities							Mid-fill blend number, % distilled at	
						RON			MON				
Stream	Specific gravity	Sulfur content, % weight	Viscosity[a]	Smoke point, mm.	R.V.P.	Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	160°F	210°F
Mixed Canadian													
Full range naphtha	.7386	.0643											
C5 to 200°F	.6845	.0422			11.7	75.3	79.8	86.3	73.9	77.6	83.3	41.0	95.0
C5 to 160°F	.6643	.0406			15.5	78.5	80.7	87.2	74.8	78.5	84.2	91.3	100.0
Light 160-200°F	.7187	.0448			9.5	73.3	78.0	84.8	72.4	76.2	81.8	0.0	99.0
Medium 200-340°F	.7619	.0529											
Heavy 340-375°F	.7911	.0817	4.53	28.8									
Atmospheric gas oil 375-650°F	.8406	.3102	15.18										
Light gas oil 375-500°F	.8239	.1720	9.04	25.6									
Heavy gas oil 500-650°F	.8682	.4382	20.88										
Atmospheric bottoms 650°F+	.9349	.8636	31.74										
Vacuum overhead 650-1050°F	.915	.7121	28.75										
Vacuum bottoms 1050°F+	1.020	1.4300	49.14										
Once through isomerate	.65	.0001			16.0	84.3	87.7	92.5	84.9	86.5	89.1	93.0	105.0
Recycle isomerate	.64	.0001			16.5	91.0	93.7	97.5	87.6	91.4	96.9	95.0	100.0
Indonesian Minas													
Full range naphtha	.7400	.0082											
C5 to 200°F	.6890	.0013			9.4	63.7	72.7	85.1	60.2	71.0	85.6	41.0	95.0
C5 to 160°F	.650	.0002			10.6	67.5	76.2	88.0	61.8	73.0	88.0	91.3	100.0
Light 160-200°F	.6882	.0024			8.13	59.7	69.2	82.0	58.5	68.5	83.0	0.0	98.0
Medium 200-340°F	.7521	.0115											
Heavy 340-375°F	.7877	.0101	2.0	31.3									
Atmospheric gas oil 375-650°F	.8134	.0255	11.43										
Light gas oil 375-500°F	.80	.0120	6.7	27.5									
Heavy gas oil 500-650°F	.823	.0349	14.8										
Atmospheric bottoms 650°F+	.8886	.1051	33.2										
Vacuum overhead 650-1050°F	.864	.0890	29.17										
Vacuum bottoms 1050°F+	.9433	.1387	42.4										
Once through isomerate	.647	.0001			11.1	78.6	84.7	93.1	76.8	83.5	92.6	93.0	105.0
Recycle isomerate	.645	.0001			11.6	87.4	91.7	97.7	83.2	90.0	95.8	95.0	100.0
Natural gasoline													
C5 to 200°F	.672	.0020			10.8	78.9	84.0	93.5	77.8	81.7	89.9	53.5	105.0
C5 to 160°F	.643	.0010			12.0	79.5	84.6	94.1	78.3	81.9	90.0	91.3	100.0
Light 160-200°F	.727	.0046			4.1	67.8	75.3	85.0	64.8	73.0	84.6	0.0	99.0
Medium 200-340°F	.782	.0093											
Once through isomerate	.637	.0001			12.5	86.2	92.4	100.8	86.5	89.6	94.2	93.0	105.0
Recycle isomerate	.630	.0001			13.0	92.0	96.3	102.8	88.3	93.1	99.8	95.0	100.0

[a] Natural blending values for viscosity in centistokes @122°F.

Table H-15. STREAM QUALITIES

Miscellaneous Streams

Stream	Specific gravity	Sulfur content, % weight	Smoke point	Gasoline blending qualities							Mid-fill blend number, % distilled at	
				R.V.P.	RON			MON				
					Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	150°F	210°F
Chemical compounds[a]												
Propane	.508	negl.										
Isobutane	.563	negl.		76.0	100.5	104.4	108.0	95.8	101.3	106.3	115.0	110.0
Normal butane	.584	negl.		59.0	92.0	96.5	103.2	89.0	94.4	102.0	115.0	110.0
Process streams[b]												
90 Sev. reformate												
Light SR feed	.78	.0001		9.0	90.5	93.7	97.8	80.1	82.9	87.1	12.2	93.1
Medium SR feed	.79	.0001		5.3	90.5	93.7	97.8	80.1	82.9	87.1	2.6	17.4
Heavy SR feed	.79	.0001		1.3	90.5	93.7	97.8	80.1	82.9	87.1	-.7	.5
Hydrocracked naphtha feed	.79	.0001		6.4	90.5	93.7	97.8	80.1	82.9	87.1	4.2	22.0
Coker naphtha feed	.79	.0001		4.3	90.5	93.7	97.8	80.1	82.9	87.1	3.0	17.8
Cat. naphtha feed	.79	.0001		4.3	90.5	93.7	97.8	80.1	82.9	87.1	3.0	17.8
95 Sev. Reformate												
Light SR feed	.79	.0001		9.2	95.3	97.2	100.2	82.1	84.7	88.5	13.5	94.5
Medium SR feed	.80	.0001		5.5	95.3	97.2	100.2	82.1	84.7	88.5	3.6	18.5
Heavy SR feed	.80	.0001		1.5	95.3	97.2	100.2	82.1	84.7	88.5	-.5	1.0
Hydrocracked naphtha feed	.80	.0001		6.7	95.3	97.2	100.2	82.1	84.7	88.5	6.8	24.5
Coker naphtha feed	.80	.0001		4.5	95.3	97.2	100.2	82.1	84.7	88.5	3.6	18.5
Cat. naphtha feed	.80	.0001		4.5	95.3	97.2	100.2	82.1	84.7	88.5	3.6	18.5
100 Sev. reformate												
Light SR feed	.80	.0001		9.5	99.8	101.8	102.9	86.0	88.0	91.0	17.0	96.2
Medium SR feed	.81	.0001		5.8	99.8	101.8	102.9	86.0	88.0	91.0	6.5	20.2
Heavy SR feed	.81	.0001		1.8	99.8	101.8	102.9	86.0	88.0	91.0	2.4	2.7
Hydrocracked naphtha feed	.81	.0001		7.0	99.8	101.8	102.9	86.0	88.0	91.0	9.7	26.2
Coker naphtha feed	.81	.0001		4.8	99.8	101.8	102.9	86.0	88.0	91.0	6.5	20.2
Cat. naphtha feed	.81	.0001		4.8	99.8	101.8	102.9	86.0	88.0	91.0	6.5	20.2

Table H-15. (continued). STREAM QUALITIES

Miscellaneous Streams

Stream	Specific gravity	Sulfur content, % weight	Smoke point, mm.	Gasoline blending qualities							Mid-fill blend number, % distilled at	
					RON			MON				
				R.V.P.	Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	150°F	210°F
Process Streams (continued)												
Light hydrocrackate	.68	.0001		13.1	85.6	90.6	97.3	82.4	88.9	95.1	78.0	100.0
Heavy hydrocrackate												
SR gas oil feed	.76	.0004	30.0									
Cracked gas oil feed	.76	.0004	15.0									
Hydrocracked jet fuel	.812	.0004	26.0									
Desulfurized cat. gasoline												
Light	.706	.0003		11.0	78.0	83.4	93.1	77.0	81.6	90.7	40.0	90.4
Heavy	.793	.0004		.5	56.0	65.0	83.2	58.0	65.0	79.2	-20.0	- 8.0
Alkylate	.699	.0004		3.5	95.0	98.6	105.4	89.8	95.3	103.9	1.5	29.3
BTX raffinate												
Light SR feed (180-200°F)[b]	.692	.0001		4.2	61.6	70.8	83.4	62.6	67.1	83.6	2.0	77.0
Medium SR feed (200-340°F)	.777	.0001		1.6	87.1	91.4	96.4	78.2	81.0	88.5	0.0	0.0

[a] Well-defined chemical compounds whose qualities are constant.

[b] Streams whose qualities are a function of the processing unit rather than the feed stream (with the exception of reformate whose specific gravity varies with feed gravity).

Table H-16. STREAM QUALITIES

Variable Sulfur Streams[a]

Stream	Specific gravity	Viscosity[b]	Gasoline blending qualities							Mid-fill blend number % distilled at	
			R.V.P.	RON			MON				
				Clear	0.5 cc	3.0 cc	Clear	0.5 cc	3.0 cc	150°F	210°F
Cat. cracked gasolines											
Low sev, raw feed	.753		6.2	92.0	94.8	98.8	80.0	82.2	85.3	17.0	46.7
High sev, raw feed	.757		6.2	93.0	95.8	99.8	80.5	82.7	85.8	17.0	46.7
Low sev, treated feed	.756		6.2	93.0	95.8	99.8	80.5	82.7	85.8	17.0	46.7
High sev, treated feed	.758		6.2	93.0	95.8	99.8	80.5	82.7	85.8	17.0	46.7
Light cycle oil	.8956	16.46									
Heavy cycle oil	.9448	22.55									
Cat. feed – desulfurized	varies	28.75									
Light coker naphtha	.677		18.2	78.0	83.3	90.7	71.2	74.8	79.9	62.5	99.0
Desulfurized	.677		15.0	77.0	82.3	89.7	72.2	75.8	80.9	62.5	99.0
Medium coker naphtha	.765										
Desulfurized	.761		1.2	55.0	60.1	67.7	52.2	56.9	63.9	-10.0	1.0
Coker gas oil	.844	16.48									
Coke	–										
Visbreaker gasoline	.735		3.3	62.3	66.0	71.4	56.6	59.1	62.6	3.0	18.0
Visbreaker gas oil	.837	16.46									
Visbreaker tar	.910	22.10									

[a] Streams whose specific gravity and viscosity are unit-dependent yet whose sulfur content varies with the feed sulfur. See tables H-17 and H-18 and Appendix J for the percentage distribution of feed sulfur among output streams and a discussion of the usage of these percentages.

[b] Refutas blending values for viscosity in centistokes @ 122°F.

Table H-17. SULFUR DISTRIBUTION

Coker and Visbreaker

(% feed sulfur)

Unit	Sulfur distribution
Coker – 1050°F+ feed	
H_2S	34.4
Light gasoline	1.2
Heavy gasoline	3.4
Gas oil	30.3
Coke	30.7
Total	100.0%
Visbreaker – 1050°F+ feed	
H_2S	10.0
Gasoline	.8
Gas oil	29.2
Tar	60.0
Total	100.0%

Table H-18. SULFUR DISTRIBUTION

Catalytic Cracking

(% feed sulfur)

Case/feed	Process conversion, % LV feed[a]	Output Stream					
		H_2S	Gasoline	Light cycle oil	Heavy cycle oil	Coke[b]	Total
Case 1 – all crude VOH's							
	65	40.0	6.0	33.0	15.0	6.0	100.0%
	72.5	44.0	6.5	30.0	13.5	6.0	100.0%
	85	50.0	7.0	21.0	14.0	8.0	100.0%
	95	52.0	7.5	19.5	13.0	8.0	100.0%
Case 2							
Crude independent VOH	72.5	20.0	3.5	34.5	33.5	8.5	100.0%
Crude dependent							
Louisiana VOH	65	41.7	5.0	18.0	30.5	4.8	100.0%
	85	48.7	4.0	13.8	26.3	7.2	100.0%
West Texas Sour VOH	65	38.1	4.1	31.0	23.5	3.3	100.0%
	85	45.1	3.1	26.8	19.3	5.7	100.0%
Oklahoma VOH	65	41.7	5.0	18.0	30.5	4.8	100.0%
	85	48.7	4.0	13.8	26.3	7.2	100.0%
Calif. Wilmington VOH	65	55.4	10.1	23.7	9.8	1.0	100.0%
	85	62.4	9.1	19.5	5.6	3.4	100.0%
Calif. Ventura VOH	65	55.4	10.1	23.7	9.8	1.0	100.0%
	85	62.4	9.1	19.5	5.6	3.4	100.0%
Alaskan North Slope VOH	65	49.2	7.6	20.5	19.5	3.2	100.0%
	85	56.2	6.6	16.3	15.3	5.6	100.0%
Nigerian Forcados VOH	65	41.7	5.0	18.0	30.5	4.8	100.0%
	85	48.7	4.0	13.8	26.3	7.2	100.0%
Arabian Light VOH	65	41.7	4.4	24.1	20.3	9.5	100.0%
	85	48.7	3.4	19.9	16.1	11.9	100.0%
Venezualan Tia Juana VOH	65	38.1	4.1	31.0	23.5	3.3	100.0%
	85	45.1	3.1	26.8	19.3	5.7	100.0%
Algerian Hassi Messaoud VOH	65	41.7	5.0	18.0	30.5	4.8	100.0%
	85	48.7	4.0	13.8	26.3	7.2	100.0%
Mixed Canadian VOH	65	41.7	5.0	18.0	30.5	4.8	100.0%
	85	48.7	4.0	13.8	26.3	7.2	100.0%
Indonesian Minas VOH	65	48.8	7.4	13.9	12.4	17.5	100.0%
	85	55.8	6.4	9.7	8.2	19.9	100.0%

[a] Conversion levels vary as follows: untreated feed, low severity (65%) and high severity (85%); treated feed, low severity (72.5%) and high severity (95%).

[b] Equivalent to SO_x production.

Table H-19. ALTERNATE YIELD DATA

High and Low Severity Reforming of SR Naphtha

90 RON Severity

Stream[a]	High pressure (450 psi)					Low pressure (225 psi)				
	Louisiana	West Texas Sour	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana	Louisiana	West Texas Sour	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana
Light feed (180-200°)										
Hydrogen (MSCF)[b]	.87	.797	1.05	.435	.86	1.12	1.08	1.26	.887	1.14
Ethane/methane (FOE)	.0383	.0425	.0147	.0770	.0358	.0227	.0262	.002	.0560	.0208
Propane	.0612	.0678	.0175	.124	.0574	.0382	.043	.0035	.0864	.0366
Isobutane	.0266	.0295	.0083	.0638	.025	.0166	.0188	.001	.0374	.0154
Normal butane	.0392	.0436	.0105	.0799	.0366	.0235	.0262	.0025	.0490	.022
90 Reformate	.796	.783	.905	.666	.807	.844	.833	.933	.7320	.851
Medium feed (200-340°)										
Hydrogen (MSCF)[b]	1.02	.985	1.045	.69	.95	1.24	1.225	1.256	.983	1.205
Ethane/methane (FOE)	.0185	.0225	.0165	.052	.0266	.0055	.0095	.0028	.0337	.0134
Propane	.025	.0333	.0203	.083	.0408	.0098	.017	.0049	.0537	.0235
Isobutane	.0114	.0147	.0095	.0361	.018	.0043	.0072	.0020	.0236	.0102
Normal butane	.0158	.0213	.012	.0535	.0282	.0084	.0107	.0030	.0320	.015
90 Reformate	.886	.864	.898	.752	.845	.917	.897	.9310	.8075	.88
Heavy feed (340-375°)										
Hydrogen (MSCF)[b]	1.05	1.02	1.065	.9	.985	1.26	1.24	1.27	1.180	1.225
Ethane/methane (FOE)	.0147	.0185	.0119	.0315	.0225	.002	.0057	.0003	.0175	.0095
Propane	.0175	.0253	.0114	.0506	.0333	.0035	.0101	.0008	.0307	.017
Isobutane	.0083	.0115	.0058	.022	.0147	.001	.0043	.0002	.0133	.0072
Normal butane	.0105	.016	.0066	.0322	.0213	.0025	.0065	.0004	.0193	.0107
90 Reformate	.905	.885	.922	.822	.864	.933	.916	.948	.8620	.897

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[a] LV Fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

Table H-19 (continued). ALTERNATE YIELD DATA

High and Low severity Reforming of SR Naphtha

95 RON Severity

Stream[a]	High pressure (450 psi)					Low pressure (225 psi)				
	Louisiana	West Texas Sour	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana	Louisiana	West Texas Sour	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana
Light feed (180-200°)										
Hydrogen (MSCF)[b]	.84	.825	1.096	.445	.89	1.255	1.215	1.37	.860	1.275
Ethane/methane (FOE)	.0488	.0625	.0228	.0647	.0467	.0286	.0328	.0103	.0735	.0262
Propane	.077	.0845	.029	.146	.0727	.0498	.054	.0168	.0885	.0475
Isobutane	.0334	.0365	.0134	.063	.0316	.0206	.0227	.0075	.0413	.0193
Normal butane	.0494	.0642	.0188	.0927	.0467	.0301	.0333	.0105	.0628	.0283
95 Reformate	.76	.745	.868	.624	.788	.806	.794	.894	.6860	.812
Medium feed (200-340°)										
Hydrogen (MSCF)[b]	1.06	1.02	1.085	.72	.985	1.38	1.35	1.388	1.127	1.33
Ethane/methane (FOE)	.0275	.0325	.0247	.0614	.0372	.0128	.0155	.0110	.0430	.0186
Propane	.0375	.0488	.0322	.1014	.066	.0231	.0298	.0185	.0626	.0365
Isobutane	.0172	.0208	.0152	.0437	.0241	.0097	.0122	.0079	.0275	.0146
Normal butane	.0243	.0302	.021	.0645	.0354	.014	.0175	.0115	.0407	.0212
95 Reformate	.847	.825	.86	.712	.806	.877	.868	.8915	.7690	.841
Heavy feed (340-375°)										
Hydrogen (MSCF)[b]	1.095	1.06	1.13	.935	1.02	1.37	1.38	1.375	1.310	1.35
Ethane/methane (FOE)	.0228	.0275	.019	.043	.0325	.0103	.013	.0085	.0221	.0155
Propane	.029	.038	.0225	.0651	.0488	.0168	.0233	.012	.0433	.0298
Isobutane	.0138	.0174	.0112	.0284	.0208	.0075	.0088	.0057	.0172	.0122
Normal butane	.0188	.0245	.0148	.0419	.0303	.0105	.0142	.0085	.0251	.0175
95 Reformate	.868	.847	.886	.783	.825	.894	.876	.909	.8240	.858

[a] LV Fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

Table H-19 (continued). ALTERNATE YIELD DATA

High and Low Severity Reforming of SR Naphtha

100 RON Severity

Stream[a]	High pressure (450 psi)					Low pressure (225 psi)				
	Louisiana	West Texas Sour	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana	Louisiana	West Texas Sour	Nigerian Forcados	Arabian Light	Venezuelan Tia Juana
Light feed (160-200°)										
Hydrogen (MSCF)[b]	.89	.85	1.17	.447	.92	1.39	1.355	1.48	1.030	1.41
Ethane/methane (FOE)	.0677	.0608	.037	.0877	.056	.0394	.044	.0175	.0881	.0388
Propane	.097	.1046	.047	.166	.0928	.063	.0661	.037	.0930	.081
Isobutane	.0424	.0455	.0206	.0726	.0406	.026	.0282	.0140	.0485	.0248
Normal butane	.0621	.0667	.03	.1065	.0594	.0383	.0418	.022	.0781	.0363
100 Reformate	.716	.702	.821	.588	.723	.761	.75	.851	.6600	.767
Medium feed (200-340°)										
Hydrogen (MSCF)[b]	1.106	1.06	1.13	.74	1.02	1.475	1.47	1.479	1.273	1.46
Ethane/methane (FOE)	.0406	.0447	.0385	.0683	.0482	.0211	.0242	.0186	.0550	.0285
Propane	.066	.0658	.0605	.1215	.0743	.042	.0475	.0382	.0729	.0626
Isobutane	.0246	.0288	.022	.0530	.0326	.0142	.0181	.0148	.0333	.0203
Normal butane	.0357	.0422	.032	.0773	.0477	.0243	.0264	.0229	.0611	.0293
100 Reformate	.802	.781	.813	.671	.762	.836	.817	.8480	.7270	.798
Heavy feed (340-375°)										
Hydrogen (MSCF)[b]	1.17	1.106	1.185	.965	1.06	1.48	1.475	1.480	1.445	1.47
Ethane/methane (FOE)	.037	.0408	.0342	.053	.0446	.0176	.0212	.0153	.0325	.0242
Propane	.047	.0564	.0401	.0856	.0658	.037	.0422	.0329	.0579	.0475
Isobutane	.0206	.0247	.0174	.0374	.0288	.014	.0163	.0129	.0227	.0181
Normal butane	.03	.036	.025	.0547	.0422	.022	.0245	.021	.0328	.0264
100 Reformate	.821	.801	.836	.738	.781	.851	.834	.864	.7790	.817

[a] LV Fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

Table H-20. ALTERNATE YIELD DATA

High and Low Pressure Reforming of Conversion Naphtha

	High pressure (450 psi)				Low pressure (225 psi)			
	Heavy hydrocracked naphtha		Medium Coker naphtha	Heavy Cat. naphtha	Heavy hydrocracked naphtha		Medium Coker naphtha	Heavy Cat. naphtha
Stream[a]	Straight run gas oil feed[c]	Cracked gas oil feed			Straight run gas oil feed[c]	Cracked gas oil feed		
90 RON Severity								
Hydrogen (MSCF)[b]	1.06	1.045	1.02	1.02	1.26	1.255	1.24	1.24
Ethane/methane (FOE)	.0147	.0165	.0185	.0185	.002	.0028	.0065	.0065
Propane	.0175	.0203	.025	.025	.0035	.0049	.0098	.0098
Isobutane	.0083	.0096	.0114	.0114	.001	.0020	.0043	.0043
Normal butane	.0106	.012	.0158	.0158	.0025	.0030	.0064	.0064
90 Reformate	.9060	.898	.8880	.8880	.933	.9310	.917	.917
96 RON Severity								
Hydrogen (MSCF)[b]	1.095	1.085	1.06	1.06	1.37	1.366	1.36	1.36
Ethane/methane (FOE)	.0228	.0247	.0275	.0275	.0103	.0110	.0128	.0128
Propane	.029	.0322	.0375	.0375	.0168	.0185	.0231	.0231
Isobutane	.0138	.0152	.0172	.0172	.0075	.0079	.0097	.0097
Normal butane	.0188	.021	.0243	.0243	.0105	.0115	.014	.014
96 Reformate	.8680	.86	.8470	.8470	.894	.8915	.877	.877
100 RON Severity								
Hydrogen (MSCF)[b]	1.17	1.13	1.105	1.105	1.48	1.479	1.475	1.475
Ethane/methane (FOE)	.037	.0385	.0406	.0406	.0175	.0185	.0211	.0211
Propane	.047	.0605	.056	.056	.037	.0362	.042	.042
Isobutane	.0206	.022	.0245	.0245	.0140	.0148	.0142	.0142
Normal butane	.03	.032	.0367	.0367	.022	.0229	.0243	.0243
100 Reformate	.8210	.813	.8020	.8020	.861	.8480	.835	.835

[a] LV Fraction on feed unless otherwise noted.

[b] Effective hydrogen yield available for hydrotreating.

[c] Includes atmospheric gas oils and vacuum overhead.

Table H-21. OPERATING COST CONSUMPTIONS

Reforming

(per Bbl intake)

	Straight-run naphtha		Heavy hydrocracked naphtha				Medium coker naphtha		Heavy Cat. naphtha	
			Straight-run gas oil feed		Cracked gas oil feed					
	Clusters	G.R.[a]	Clusters	G.R.[a]	Clusters	G.R.[a]	Clusters	G.R.[a]	Clusters	G.R.[a]
90 Severity										
Maintenance ($)	.06	.06	.06	.06	.06	.06	.06	.06	.06	.06
Labor (shift positions)	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001
Catalyst and chemicals ($)	.04	.04	.04	.04	.04	.04	.04	.04	.04	.04
Electricity (KWH)	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Steam (Mlbs)	.001	.001	.001	.001	.001	.001	.001	.001	.001	.001
Cooling water (Mgal)	.6	.45	.6	.45	.6	.45	.6	.45	.6	.45
Refinery fuel (FOE)	.05	.05	.05	.05	.05	.05	.05	.05	.05	.05
95 Severity										
Maintenance ($)	.062	.062	.062	.062	.062	.062	.062	.062	.062	.062
Labor (shift positions)	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001
Catalyst and chemicals ($)	.06	.06	.06	.06	.06	.06	.06	.06	.06	.06
Electricity (KWH)	3.2	3.2	3.2	3.2	3.2	3.2	3.2	3.2	3.2	3.2
Steam (Mlbs)	.001	.001	.001	.001	.001	.001	.001	.001	.001	.001
Cooling water (Mgal)	.7	.5	.7	.5	.7	.5	.7	.5	.7	.5
Refinery fuel (FOE)	.05	.05	.05	.05	.05	.05	.05	.05	.05	.05
100 Severity										
Maintenance ($)	.064	.064	.064	.064	.064	.064	.064	.064	.064	.064
Labor (shift positions)	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001	.0001
Catalyst and chemicals ($)	.10	.10	.10	.10	.10	.10	.10	.10	.10	.10
Electricity (KWH)	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5
Steam (Mlbs)	.001	.001	.001	.001	.001	.001	.001	.001	.001	.001
Cooling water (Mgal)	.8	.55	.8	.55	.8	.55	.8	.55	.8	.55
Refinery fuel (FOE)	.05	.05	.05	.05	.05	.05	.05	.05	.05	.05

[a] Grassroots.

Table H-22. OPERATING COST CONSUMPTIONS

Catalytic Cracking

(per Bbl intake)

	Hydrotreated feed			Raw feed			
	Low severity		High severity[a]	Low severity		High severity	
	Clusters	Grassroots	Grassroots	Clusters	Grassroots	Clusters	Grassroots
Maintenance ($)	.049	.049	.051	.049	.049	.051	.051
Labor (shift positions)	.00015	.00015	.00015	.00015	.00015	.00015	.00015
Catalyst & chemicals ($)	.04	.04	.08	.04	.04	.08	.08
Electricity (KWH)	2.0	2.0	2.4	2.0	2.0	2.4	2.4
Steam (Mlbs)	.001	.001	.001	.001	.001	.001	.001
Cooling water (Mgal)	1.0	.8	1.0	1.0	.8	1.2	1.0
Refinery fuel (FOE)	.024	.024	.048	.02	.02	.04	.04

[a] High severity catalytic cracking of hydrotreated feed is not used in the cluster models.

Table H-23 OPERATING COST CONSUMPTIONS

Hydrocracking

(Per Bbl intake)

	High severity hydrocracking									Medium severity hydrocracking	
	Heavy gas oil feed			Vacuum gas oil feed			Cracked gas oil feed			All gas oil feeds	
	Clusters	East Grassroots (sour crude)	West and East Grassroots (sweet crude)	Clusters	East Grassroots (sour crude)	West and East Grassroots (sweet crude)	Clusters	East Grassroots (sour crude)	West and East Grassroots (sweet crude)	Clusters	Grassroots
Maintenance ($)	.075	.079	.075	.075	.079	.075	.075	.079	.075	.077	.077
Labor (shift positions)	.00008	.00008	.00008	.00008	.00008	.00008	.00008	.00008	.00008	.00008	.00008
Catalyst & chemicals ($)	.10	.105	.10	.10	.105	.10	.10	.125	.10	.10	.10
Electricity (KWH)	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	9.0	9.0
Steam (Mlbs)	.015	.015	.015	.015	.015	.015	.015	.015	.015	.0145	.0145
Cooling water (Mgal)	.7	.5	.5	.7	.5	.5	.7	.5	.5	.68	.48
Refinery Fuel (FOE)	.03	.03	.03	.03	.03	.03	.03	.03	.03	.028	.028

Table II-24. OPERATING COST CONSUMPTIONS

Desulfurization[a]

(per Bbl intake)

	5/" Naphtha	Light + heavy gas oil	Vacuum overhead	Cat. naphtha	Light + medium coker naphtha	Light cycle oil	Atm Btms to .3% North Slope	Atm Btms to 1% Arabian Light	Atm Btms to .5% Arabian Light	Vac Btms to .8% North Slope	Vac Btms to 1% Arabian Light	Vac Btms to .8% Arabian Light
Maintenance ($)	.014	.022	.025	.014	.014	.023	.048	.06	.064	.07	.07	.076
Labor (shift positions)	.0001	.00013	.00008	0001	.0001	.00013	.000075	.000075	.000075	.0002	.0002	.0002
Catalyst + Chemicals ($)	.013	.015	.02	.015	.015	.017	.063	.085	.085	.13	.13	.17
Electricity (KWH)	1.0	1.2	1.5	1.0	1.0	1.3	4.7	6.2	6.3	8.7	8.7	9.5
Steam (Mlbs)	.01	.01	.01	.01	.01	.01	.01	.01	.01	.02	.02	.02
Cooling water (Mgal)	.3	.4	.4	.3	.3	.4	.14	.17	.18	.22	.22	.25
Refinery fuel (FOE)	.01	.01	.01	.01	.01	.01	.012	.013	.013	.012	.012	.012

[a] All cluster and grass roots models.

Table H-25. OPERATING COST CONSUMPTIONS

Miscellaneous Process Units[a]

(per Bbl intake)

	Sulfur recovery units[b]		Isomerization		H_2 Manufacture[c]	
	95% Recovery	99.95% Recovery	Once thru	Recycle	Clusters	Grassroots
Maintenance ($)	.56	1.06	.03	.06	0.032	0.032
Labor (shift positions)	.028	.028	.0002	.0002	0.00014	0.00014
Catalyst and chemicals ($)	6.0	7.6	.085	.12	0.005	0.005
Electricity (KWH)	80.0	346.0	2.4	4.8	0.17	0.17
Steam (Mlbs)	4.0	3.93	.01	.01	0.001	0.001
Cooling water (Mgal)	.4	.48	.3	.3	0.1	0.1
Refinery fuel (FOE)	2.4	2.9	.03	.03	0.031	0.031
Make-up water					—	0.011

	Coking	Visbreaking	Alkylation[d]	Aromatics extraction	Atmospheric distillation	Vacuum distillation	Crude and products handling
Maintenance ($)	.086	.0231	.084	.05	.008	.009	.032
Labor (shift positions)	.0002	.0002	.0003	.004	.00002	.00005	.00008
Catalyst and chemicals ($)	.002	.0021	.2	.132	.001	.002	.001
Electricity (KWH)	1.5	1.0	3.7	2.0	.25	.20	1.0
Steam (Mlbs)	.03	.02	.03	.4	.023	.03	.02
Cooling water (Mgal)	.6	.5	2.0	.4	.5	.20	.2
Refinery fuel (FOE)	.03	.03	.1	.005	.01	.01	.01

[a] All models.

[b] Consumption per ton of sulfur recovered. Includes associated costs required to achieve specified recovery level.

[c] Consumption per MSCF of hydrogen product.

[d] Consumption per Bbl product.

Table H-26. OPERATING COSTS COEFFICIENTS

Regional Supply Cost per Unit Coefficient

($ per unit)

	Maintenance	Labor (shift positions)	Electricity (KWH)	Catalyst and chemicals	Cooling water[a] (M gallons)	Make-up water[b] (M gallons)
Louisiana Gulf	1.32	473.0	.0144	1.1	.044	
Texas Gulf	1.32	473.0	.0144	1.1	.044	
West Coast	1.32	473.0	.018	1.1	.044	
Large Midwest	1.39	473.0	.018	1.1	.044	
Small Midcontinent	1.39	473.0	.018	1.1	.044	
East Coast	1.44	520.3	.0215	1.21	.048	
West Grassroots	1.38	500.0	.02	1.15		.35
East Grassroots	1.38	500.0	.02	1.15		.35

[a] Cluster models purchase cooling water.

[b] Grassroots models generate cooling water in a cooling tower which consumes purchased make-up water to compensate for evaporation losses.

Table H-27. PROCESS UNIT CAPITAL INVESTMENT ESTIMATES

Process unit	Size basis, MB/SD	Investment estimate $/B/SD
Atmospheric distillation	100	165
Vacuum distillation	40	185
Catalytic cracking	40	925
Catalytic reforming (low pressure)	20	800
Alkylation (product basis)	10	1400
Isomerization – once through	10	620
Isomerization – recycle	10	1240
Hydrocracking (high severity)	25	1400
Naphtha hydrotreating	20	235
FCC/coker gasoline hydrotreating	15	320
Light distillate hydrotreating	30	230
Heavy distillate hydrotreating	30	250
Vacuum gas oil desulfurization (FCC feed)	25	370
Atmospheric residual desulfurization	50	775
Vacuum residual desulfurization	15	1500
Coking – delayed	10	930
Hydrogen generation-methane MMSCF/SD	50	230[a]
-naphtha MMSCF/SD	50	260[a]
Sulfur recovery (95% removal) Short Tons/SD	100	25000
Sulfur recovery (99.95% removal) Short Tons/SD	100	50000

[a] $/MSCF/SD

Table II-28. OFFSITE AND OTHER ASSOCIATED COSTS OF REFINERIES USED IN ESTIMATING COST OF GRASSROOTS REFINERIES

1st Quarter 1975 Basis

(% onsite cost)

Type of cost		Refinery complexity							
		3	4	5	6	7	8	9	10
Mainly complexity-related offsites, % — Includes utilities, piping, blending, product handling, buildings, roads, site preparation, safety and fire.		138.0	103.0	88.9	74.7	70.4	64.9	61.7	57.4
Other offsites, % — Includes tankage, ecology and land		87.0	67.0	59.0	51.0	48.0	44.2	42.0	39.0
Offsites—subtotal	%	225.0	170.0	147.9	125.7	118.4	109.1	103.7	96.4
Associated costs									
Chemicals and catalysts		6.0	5.0	4.5	4.0	3.8	3.5	3.3	3.0
Marine or equivalent facilities		20.0	15.5	12.8	10.0	8.8	7.8	6.8	5.8
Working capital		←	←	←	70.0	→	→	→	→
Other — Includes training, spares, telephone, autos, domestic water, cafeteria and recreation.		←	←	←	20.0	→	→	→	→
Associated—subtotal	%	116.0	110.5	107.3	104.0	102.6	101.3	100.1	98.8

Technical Documentation Reference List

1. Alcocak, L., et al., "BP Hydrocracks for Middistillates". Oil and Gas Journal July, 1974, pp. 102-110.

2. Anderson, R.F., "Changes Keep HF Alkylation Up-To-Date", Oil and Gas Journal, February, 1974, pp. 78-81.

3. "BP Adds Hydrocracking To It's Lavera Refinery", Petroleum & Petrochemical International, November, 1973, pp. 66-69.

4. Bernstein, J.L. Dauber, "Flexibility and Capability of Powerforming Are Extended", Oil and Gas Journal, September, 1968, pp. 163-167.

5. Blazek, James, "Zeolitic Catalyst Prooved On Natural Stock", Oil and Gas Journal, November, 1971, pp. 66-73.

6. Bour, George, C.P. Schwoerer, G.F. Asselin, "Penex Unit Peps Up SR Gasoline", Oil and Gas Journal, October, 1970, pp. 57-61.

7. Dupont, (letter from), to Mr. Marshall Nicols, Re: NPC study on factors affecting U.S. petroleum refining, October, 1974.

8. Dupont, (letter from), to Mr. W.A. Johnson, Re: Motor octane blending values for reformate, February, 1975.

9. Ethyl Corporation, (letter from), to Gilbert H. Wood, Re: Simulation studies for E.P.A., gasoline component blending values, December, 1974.

10. Gould, G.D., C.S. McCoy, "Rheniforming Scores High In Commercial Runs", Oil and Gas Journal, November, 1970, pp. 49-53.

11. Hainsselin, M.H., M.F. Symoniak, G.R. Cann, "Strategy of Isomerization," Hydrocarbon Processing, April, 1975, pp. 62-H - 62-T.

12. Huling, G.P., et al., "Feed-Sulfur Distribution in FCC Product", Oil and Gas Journal, May, 1975, pp. 73-79.

13. "Hysomer May Ease Lead Elimination", Oil and Gas Journal, March, 1971, pp. 44-45.

14. Jones, H.B., "Modern Cat Cracking for Smaller Refineries", Oil and Gas Journal, December, 1969, pp. 50-53.

15. Magee, et al., "Catalyst's Developments in Catalytic Cracking" Oil and Gas Journal, July, 1973, pp. 48-54.

16. Marathon Oil Company, (letter from), to Mr. Marshall W. Nicols, Re: Comments as to ADL's octane data and gasoline blending, September, 1974.

17. Murphy, J.P., M.R. Smith, C.H. Vienes, "Hydrocracking Vs. Cat Cracking for Gas Oils in Today's Refinery", Oil and Gas Journal, June, 1970, pp. 108-112.

18. "NPRA Panel", Hydrocarbon Processing, March, 1971, pp. 96-98.

19. "NPRA Panel", Oil and Gas Journal, February, 1971, p. 63.

20. Nelson, W.L., "Costs of Alkylation and Viscosity-Breaking Plants are Updated", Oil and Gas Journal, April, 1974, pp. 74-75.

21. ------, "Cost of Catalytic-Cracking Plants", Oil and Gas Journal, April, 1974, pp. 66-67.

22. ------, "Cost Examination for Coking Plants", Oil and Gas Journal, April, 1974, pp. 118-119.

23. ------, "Cost of Refineries - Part 1: Offsite Facilities", Oil and Journal, July, 1974, pp. 114-116.

24. ------, "Cost of Refineries - Part 2: Process-Unit Costs", Oil and Gas Journal, July, 1974, p. 87

25. ------, "Cost of Refineries - Part 3: Off-Sites Breakup", Oil and Gas Journal, July, 1974, pp. 60-61.

26. ------, "Cost of Refineries - Part 4: Storage, Environment, Land", Oil and Gas Journal, July, 1974, pp. 160-162.

27. ------, "Hydrocracking, Hydrogen-Manufacture Costs", Oil and Gas Journal, March, 1974, pp. 120-124.

28. ------, "Hydrogen Consumption In Treating", Oil and Gas Journal, January, 1972, p. 67.

29. ------, "How Much Hydrogen Is Consumed in Treating", Oil and Gas Journal, December, 1970, pp. .

30. ------, "A Look at Catalytic Reforming", Oil and Gas Journal, April, 1974, pp. .

31. ------, "A Look At Sulfur-Recovery Costs", Oil and Gas Journal, March, 1974, pp. 120-123.

32. ------," A Look at Vacuum-Distillation Costs", Oil and Gas Journal, March, 1974. p. 100.

33. ------, "Plant Costs For Processing Hydrogen", Oil and Gas Journal, March, 1974, pp. 111-112.

34. ------, "What Are Coking Yields", Oil and Gas Journal, January, 1974, p. 70.

35. ------, "What's Happening To Refinery-Construction Costs?", Oil and Gas Journal, February, 1974, pp. 70-71.

36. 'Petroleum Coke Takes On New Luster", Oil and Gas Journal, September, 1970, pp. 73-76.

37. Production of Low Sulfur Gasoline, Contract# 68-02-1308, Task 10, Phases 1, 2, 3, can be obtained by writing to M.W. Kellogg Company, 1300 Tree Greenway Plaza East, Houston, Texas, 77046.

38. "1974 Refining Process Handbook", *Hydrocarbon Processing*, September, 1974, pp. 107-113, 116-119, 128-131, 207-208, 211-213.

39. Rose, K.E., "Delayed Coking - What You Should Know", *Hydrocarbon Processing*, July, 1971, pp. 85-92.

40. Stockey, A. Nelson, Richard F. Bauman, "Cyclic Powerforming Ups Octane", *Hydrocarbon Processing*, May 1971, pp. 106-110.

41. Texaco, (letter from), to Mr. George Holzman, Re: Gasoline blending data used by Arthur D. Little for various lead studies for the E.P.A., November, 1974.

42. "Two New Intermediate Activity Catalysts Developed", *Oil and Gas Journal*, October, 1971, pp. 82-83.

43. *U.S. Motor Gasoline Economics, Volume 1, Manufacture of Unleaded Gasoline*, American Petroleum Institute,1967.

44. Wachtel, S.J., et al., "Atlantic Richfields Lab Unit Apes Fluid Catalytic Cracker", *Oil and Gas Journal*, April, 1972, pp. 104-107.

45. Ward, J.W., A.D. Reichle, J. Sosnowski, "Catalyst Advance Open Doors for Hydrocracking", *Oil and Gas Journal*, May, 1973, pp. 69-73.

46. Whillington, E.L., Murphy, I.H. Lutz, "Striking Advances Show Up In Modern FCC Design", *Oil and Gas Journal*, October, 1972, pp. 49-54.

47. White, Paul, J., "How Cracker Feed Influences Yield", *Hydrocarbon Processing*, May, 1968, pp. 103-108.

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APPENDIX I

MODEL CALIBRATION

Upon completion of the development of the cluster refinery modeling concept, which is discussed in Appendix F, an extensive calibration effort was undertaken by Arthur D. Little (ADL) with the assistance of the Bureau of Mines (BOM), Environmental Protection Agency (EPA), and an ad hoc industry task force coordinated by the American Petroleum Institute (API) and National Petroleum Refiners Association (NPRA).

A. BASIC DATA FOR CALIBRATION

1. Refinery Input/Output

Every refiner in the United States provides monthly statistics to the BOM concerning refinery inputs, production and a summary of fuel consumed for all purposes. The BOM accumulates and summarizes this data on a monthly and annual basis, and publishes aggregate statistics by PAD district, BOM refining district, and in some cases by state.

For this project, the BOM supplied EPA with 1973 annual data for the aggregate of the three specific refineries comprising each individual cluster model (see Appendix F). This data as received from the BOM is presented in Tables I-1, I-2, and I-3.

Table I-1 contains the refinery input/output data for the year 1973 in the standard BOM reporting format.

Table I-2 provides a breakdown of refinery crude receipts by origin, by individual state for domestic crudes, and by country for foreign sources. Tables I-3 and I-4 provide statistics on fuel consumed for all purposes in the cluster model refineries for the year 1973.

Table I-1. BUREAU OF MINES REFINERY INPUT/OUTPUT DATA FOR CLUSTER MODELS: 1973

(Mbbls)

	Louisiana Gulf		Texas Gulf		Small Midcontinent		Large Midwest		West Coast		East Coast	
	Input	Output	Input	Output	Input	Output	Input	Output	Input	Output	Input	Output
1. Crude oil (including lease condensate)												
a. Domestic	238,190		317,831		49,988		128,062		106,067		47,340	
b. Foreign	2,112		40,862		10,733		30,247		69,507		157,890	
2. Products of natural gas processing plants												
a. Propane	2,109		–		–		–		2		–	
b. Isobutane	6,885		2,334		1,028		4,090		570		383	
c. Normal butane	5,875		2,246		340		–		93		1,927	
d. Other butanes	–		–		–		–		124		–	
e. Butane-propane mixtures	664		–		–		–		47		–	
f. Natural gasoline and isopentane	4,083		13,840		4,788		108		1,422		83	
g. Plant condensate	–		16,082		1,226		915		–		206	
3. Other hydrocarbons and hydrogen consumed as raw materials	–		31		–		–		935		432	
4. Unfinished oils	3,061	7,380	4,838	12,242	1,488	768	[illegible]	3,404	8,257	2,198	25,033ᵃ	
5. Gasoline												
a. Motor		130,006		178,081		40,891		88,467		72,734		114,904
b. Aviation		80		3,270		–		–		1,933		–
6. Special naphthas (solvents)		–		7,231		7		2,248		1,380		13
7. Jet fuel												
a. Naphtha-type		867		2,008		472		143		3,584		1,730
b. Kerosine-type		30,295		25,115		911		2,297		21,050		5,956
8. Kerosine (including range oil)		6,021		8,429		38		1,813		194		3,711
9. Distillate fuel oil		69,914		88,491		17,374		44,678		23,801		40,818
10. Residual fuel oil		5,856		17,170		262		8,084		33,457		14,053
11. Lubricating oils												
a. Bright stock		–		1,470		111		–		84		323
b. Neutral		(12)		3,500		84		–		165		1,488
c. Other		–		12,637		175		–		152		3,263
12. Asphalt		1,701		1,536		1,876		2,813		2,199		10,856
13. Wax												
a. Microcrystalline		–		30		–		–		–		44
b. Crystalline-fully refined		–		236		–		–		–		289
c. Crystalline-other		–		264		–		–		–		–
14. Petroleum coke												
a. Marketable		4,512		6,380		1,430		4,026		10,486		–
b. Catalyst		1,880		4,072		1,184		3,301		2,630		2,528
15. Road oil		–		–		–		2,154		16		–
16. Still gas												
a. Petrochemical feedstock use		–		776		70		–		–		882
b. Refinery gas		10,463		11,657		1,874		5,860		10,254		7,427
17. Ethane and/or ethylene: Petrochemical feedstock use		667		1,268		520		–		508		58
18. Propane and/or propylene												
a. Petrochemical feedstock use		2,282		3,719		853		–		1,131		3,083
b. Other use		8,844		4,198		1,578		3,483		1,658		7,380
19. Butane and/or butylene												
a. Petrochemical feedstock use		333		631		–		–		–		1,441
b. Other use		–		1,272		8		6		1,046		–
20. Butane-propane mixtures												
a. Petrochemical feedstock use		38		–		–		–		280		–
b. Other use		–		–		1		–		1,082		–
21. Isobutane (C_4): Petrochemical feedstock use		–		–		–		–		103		–
22. Naphtha less than 400° end point: petrochemical feedstock use		43		6,623		1,536		1,005		4,271		1,485
23. Other oils – over 400° end point: petrochemical feedstock use		3,464		2,864		185		–		368		29
24. Other finished products		1,467		1,760		84		427		611	731	1,213
25. Overage (input) or shortage (output)	10,823	79	8,788	1,098	2,867	132	5,850	46	11,273	380	6,965	–
26. Total	275,711	275,711	407,560	407,560	72,510	72,510	174,543	174,543	198,331	198,333	240,970	240,970

ᵃIncludes the following unfinished oils: alkylation feed, 203; reformer feed, 3,024; cat. cracker feed, 12,486; stock wax, 882; slop oil, 340; mineral oil, 386; hydrogen, 13; polymerization feed, 651; toluene, 524; alkylate, 105; and naphtha, 5,700 for a total of 25,033.

Table I-2. BUREAU OF MINES RECEIPTS OF CRUDE BY ORIGIN 1973

(Mbbls)

	Louisiana Gulf	Texas Gulf	Small Midcontinent	Large Midwest	West Coast	East Coast
Domestic crudes						
State of origin:						
Alabama	50	5,231				
Alaska					12,146	
California					89,254	
Colorado			459	3,398		
Florida		12,051				2,594
Illinois			63	9,090		
Kansas			18,906	836		
Louisiana	185,654	18,306	2,259	19,354		3,346
Oklahoma	2,395	1,601	21,931	10,643		
Mississippi						275
Montana				241		
Nebraska			85	836		
New Mexico			370	32,673	678	
Texas	40,502	281,252	4,686	48,805		37,800
Utah					4,321	
Wyoming			1,022	10,818		
Total domestic crudes	**228,601**	**318,441**	**49,781**	**136,694**	**106,399**	**44,015**
Foreign crudes						
Country of origin:						
Algeria						25,248
Angola		3,869				
Canada			10,744	26,022		
Ecuador					7,019	
Indonesia	161					1,165
Iran		3,666			5,920	3,905
Iraq					515	
Libya	214	50				16,121
Mexico		489				
Nigeria	827	10,213		238		29,430
Qatar					2	
Saudi Arabia	910	15,732		4,291	27,056	6,036
Sumatra					24,712	
Trinidad	189					1,772
Tunisia						3,733
United Arab Emirates	546				3,927	5,676
Venezuela	263	7,257			1,295	61,344
Total foreign crudes	**3,110**	**41,276**	**10,744**	**30,551**	**70,446**	**154,430**
Total crude	**231,711**	**359,717**	**60,525**	**167,245**	**176,845**	**198,445**

Table I-3. BUREAU OF MINES REFINERY FUEL CONSUMPTION FOR CLUSTER MODELS 1973[a]

Commodity	Unit of measure[b]	Louisiana Gulf	Texas Gulf	Small Midcontinent	Large Midwest	West Coast	East Coast
A. Fuel (purchased and produced at refinery):							
1) Crude oil used as fuel	Bbls	–	–	–	–	–	–
2) Fuel oils:							
a) Distillate-type	Bbls	330,495	345,882	198	–	209,461	–
b) Residual-type (incl. acid sludge)	Bbls	1,207,201	–	598,458	2,741,952	576,702	4,996,043
3) Liquified petroleum gases	Bbls	1,549,390	127,940	35,797	134,196	1,079,793	–
4) Natural gas	MCF	37,183,993	201,687,751	14,116,609	1,677,499	23,388,789	16,015,802
5) Still gas	MCF	65,393,750	72,856,250	12,337,500	36,625,000	64,087,500	48,125,000
6) Petroleum coke:							
a) Marketable	Short tons	–	–	–	–	–	59,846
b) Catalyst	Short tons	375,934	994,400	236,800	660,200	526,000	395,000
7) Coal	Short tons	–	–	–	–	–	–
B. Electrical energy:							
1) Purchased	MKWH	657,644	687,585	310,680	803,985	1,381,360	1,449,291
2) Generated	MKWH	5,627	502,218	–	118,820	–	46,460
3) Sold	MKWH	–	8,631	–	–	–	84,068
C. Steam:							
1) Purchased	MLbs	560,743	–	–	–	–	6,316,090
2) Sold	MLbs	28,017	–	–	–	87,984	38,302

[a] Includes consumption for three refineries for the calendar year 1973.

[b] See table I-4 for FOE conversion factors.

Table I-4. BUREAU OF MINES REFINERY FUEL CONSUMPTION FOR CLUSTER MODELS 1973[a]

(Bbls FOE)

Commodity	Conversion factor, FOE bbl per unit[b]	Louisiana Gulf	Texas Gulf	Small Midcontinent	Large Midwest	West Coast	East Coast
A. Fuel (purchased and produced at refinery):							
(1) Crude oil used as fuel	–	–	–	–	–	–	–
(2) Fuel oils:							
a) Distillate-type	.9246 per bbl	305,576	319,802	183	–	193,668	–
b) Residual-type (incl. acid sludge)	.9979 per bbl	1,204,666	–	597,201	2,736,194	575,491	4,985,551
(3) Liquified petroleum gases	.6367 per bbl	986,497	81,459	22,792	85,443	687,504	–
(4) Natural gas	.1683 per MCF	6,258,066	33,944,049	2,375,825	282,323	3,936,333	2,695,459
(5) Still gas	.1638 per MCF	10,711,496	11,933,853	2,020,883	5,999,175	10,497,532	7,882,875
(6) Petroleum coke:							
a) Marketable	4.7810 per short ton	–	–	–	–	–	286,124
b) Catalyst		1,797,340	4,754,226	1,132,141	3,156,416	2,514,806	1,888,495
(7) Coal	–	–	–	–	–	–	–
B. Electrical energy:							
(1) Purchased	1.6475 per MKWH	1,083,468	1,132,796	511,845	1,324,565	2,275,791	2,387,707
(2) Generated		9,270	827,404	–	195,756	–	76,543
(3) Sold		–	14,220	–	–	–	138,502

[a] Includes consumption for three refineries for the year 1973.

[b] One FOE (fuel oil equivalent) barrel contains 6.3×10^6 BTU's (gross heating value).

Several revisions were made to the data as originally transmitted by the BOM in preparing Table I-1. These changes resulted from discussions between the EPA and individual refiners and revised information was transmitted to ADL. Data presented in Table I-1 represents the revised version.

One of the refineries included in the East Coast cluster model shut down a catalytic cracking/alkylation complex during the calendar year 1973 and started up a new hydrocracker and associated operations. Therefore, it was considered that the use of annual statistics for this refinery in 1973 would not provide meaningful information. Accordingly, this particular plant provided specific refinery input/output data to the EPA covering that portion of the year when operations were relatively consistent. The EPA extrapolated these results to a calendar year basis which were then incorporated with the 1973 annual operating data supplied by the other companies to obtain the revised input/output data for this cluster model, and presented in Table I-1. Note that this cluster model also provided an extensive breakdown of unfinished oil intake which represents a substantial portion of the raw material consumed in this cluster.

The product blending and input streams used in the ADL refinery simulation model system do not correspond exactly to the format used in the Bureau of Mines reports. Accordingly, it was necessary to make some adjustments to the basic Bureau of Mines data to meet the requirements of the ADL model. The transformation of this data is summarized in Table I-5, with the methodology outlined in Table I-6. In comparing the data in Table I-1 with Table I-5, it must be remembered that the BOM presents annual aggregate statistics for the total of three plants while the information in Table I-5 is presented in MB/CD for the "average" single refinery operation representing the cluster. In undertaking the calibration effort, all product demands and non-crude inputs to the model were fixed. The model was then allowed to vary crude intake as needed to balance refinery product (and internal fuel) demands.

Table I-5. ADL MODEL INPUT/OUTTURN DATA FOR CALIBRATION

(MB/CD)[a]

Specified product outturns	**Louisiana Gulf**	**Texas Gulf**	**Large Midwest**	**Small Midcontinent**	**East Coast**	**West Coast**
Refinery gas/ethane (FOE)	0.32	.84	–	0.54	0.86	0.46
LPG-fuel	5.97	4.96	3.19	1.45	6.73	3.99
LPG-petrochemicals	2.40	3.97	–	0.60	4.13	1.30
Gasoline	118.79	165.02	81.70	37.44	104.93	67.77
Naphtha	0.78	8.80	2.16	0.35	1.28	3.81
BTX	–	6.05	0.94	1.40	1.36	3.90
Jet fuel	18.53	23.49	2.12	0.92	5.76	19.89
Kerosene	5.50	7.70	1.66	0.04	3.39	0.17
Distillate fuel oil	67.80	83.27	40.80	16.04	45.52	22.15
Lube stocks	–	16.49	–	0.34	4.94	0.35
Residual fuel oil	5.35	15.68	7.39	0.23	12.83	30.55
Asphalt	1.55	1.40	3.81	1.81	18.13	2.02
Coke	4.12	4.00	3.68	1.31	–	9.58
Cat. cracker feed	2.161	2.00				
Cat. reformer feed	1.164	4.955				
Specified inputs						
Isobutane	6.10	2.13	3.70	0.94	0.35	0.50
Normal butane	5.97	2.05	–	0.31	1.76	0.16
Natural gasoline	4.28	16.00	0.93	5.50	5.84	1.30
Max natural gas (FOE)	5.40	29.24	0.24	2.046	2.50	6.39
Cat. cracker feed			1.215	.436	11.10	3.597
Cat. reformer feed			.654	.235	5.98	1.937

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[a] Obtained by dividing BOM data by 365 days/year and 3 refineries per cluster.

Table I-6. CONVERSION OF BOM INPUT/OUTTURN DATA TO ADL MODEL FORMAT

ADL product/input category (as shown in Table 1-5)	Bureau of Mines category (as shown in Table I-1)
Refinery gas/ethane (FOE)	16(a) + 17 (converted to FOE basis)
LPG - fuel	18(b) + 19(b) + 20(b) - 2(a)
LPG-petrochemicals	18(a) + 19(a) + 20(a)
Gasoline	5(a) + 5(b)
Naphtha	6 + 80% of 7(a)
BTX	22
Jet fuel	20% of 7(a) + 7(b)
Kerosene	8
Distillate fuel oil	9 + 23
Lube stocks	11(a,b,c) + 13(a,b,c)
Residual fuel oil	10
Asphalt	12 + 15
Coke	14(a)
Cat cracker feed	65% of 4 (net of input/production)
Cat reformer feed	35% of 4 (net of input/production)
Isobutane	2(b)
Normal butane	2(c)
Natural gasoline	2(f) + 2(g)

Table I-6 presents the procedure for converting the BOM data as shown in Table I-1 to the ADL format in Table I-5. For the Small Midcontinent, East Coast, and West Coast cluster models, the ethane and/or ethylene production shown in line 17 of the BOM data was not converted to an FOE basis. For the West Coast cluster model, the total gasoline outturn used for the calibration model was 67.77 MB/CD although the correct number should have been 68.19. Isobutane and normal butane for Louisiana and isobutane for the West Coast were entered as 6.10/5.97/.50, respectively, instead of the correct figures 6.40/5.67/.60. In addition, LPG to petrochemicals in the West Coast was entered as 1.30 rather than the correct 1.39. Since it was felt that the calibration would not significantly be improved by correcting these items, no further adjustments were made.

The following assumptions were used in constructing Table I-6. Naphtha jet fuel production (BOM category 7 (a)) was assumed to consist of 80% naphtha and 20% kerosene. Category 22 (Naphtha-less than 400°F end point-petrochemical feedstock use) was considered to be 100% mixed aromatics referred to as BTX. Discussions with some individual oil companies indicated that for most companies this is a reasonable assumption. However, for certain companies this category represents a mixed reformate stream prior to extraction. Category 23 (other oils--over 400° end point-petrochemical feedstock use) was added to distillate fuel oil production.

Unfinished oils were considered to consist of 65% catalytic cracker feed and 35% catalytic reformer feed, except in the Texas Gulf cluster. Discussions with members of the API/NPRA Task Force indicated this to be a reasonable representation for this category. In the Texas Gulf cluster the percentage of catalytic cracker feed in unfinished oils was adjusted to 29% to improve calibration.

Discrepancies were noted in the mode of reporting BOM statistics from company-to-company. Refinery residual fuel production (BOM category 10 in Table I-1) was usually reported as the net production from the refinery. However, some companies also included in this total any internal residual fuel consumption (category A., 2b), as is shown in Table I-3. Table I-3 also provides purchased natural gas consumption in category A.4. Most companies report natural gas in this category for fuel use only. However, some companies with hydrogen generating facilities include natural gas for this purpose within this BOM category. The above company-to-company variations obviously limit the degree of calibration possible with the cluster models.

Table I-7 reports the crude slate used for each cluster model to simulate the reported BOM data. The objective was to simulate average domestic/foreign mix, sulfur content, API gravity, and other key properties as closely as possible, while still keeping the number of crudes to a manageable level. Table I-7 also contains a comparison between the average sulfur content of the model crude slates compared with the industry data obtained from the individual companies and averaged by the EPA.

The following general methodology was used to develop the model crude slates. Louisiana crude was used to simulate Louisiana and low sulfur Texas crudes. Oklahoma crude was used to represent light, sweet crudes from the Midcontinent. West Texas sour was used to simulate high sulfur crudes from Texas and New Mexico. Wilmington and Ventura were used to simulate heavy and light California crudes respectively. Nigerian Forcados was used to represent heavy African crudes while Algerian Hassi Messaoud was used to simulate light African crudes. Arabian Light represented average Middle East production and Tia Juana Medium was used to simulate Venezuelan crude.

2. Processing Configurations

The annual refinery surveys published in the Oil and Gas Journal were used as the basic reference source for determining cluster model processing configurations. Since operations for the calendar year 1973 were to be simulated, unit capacities were tabulated for January 1, 1973 and January 1, 1974, and an arithmetic average of these was used as the capacity available for the calendar year 1973. Tables I-8 through I-13 provide the basic processing data for each of the refineries comprising the respective cluster models. Each table provides 1973 and 1974 capacity data and the arithmetic average for the cluster of each, as well as the final capacity limits used in the calibration effort. It should be noted that the Oil and Gas Journal processing unit capacity data is presented in barrels per stream day; these figures were used directly for the

Table I-7. ADL MODEL CRUDE SLATES AND SULFUR CONTENTS FOR REFINERY CLUSTERS

Crude charge % total volume	Louisiana Gulf	Texas Gulf	Small Midcontinent	Large Midwest	West Coast	East Coast
Domestic crudes						
Louisiana	88.7	47.4	7.6	6.0		15.4
West Texas Sour	11.3	41.4	13.1	70.3		7.6
Oklahoma			61.5	4.9		
California Wilmington					37.4	
California Ventura					13.8	
Subtotal domestic crudes	100.0	88.8	82.2	81.2	51.2	23.0
Foreign crudes						
Nigerian Forcados		3.8				16.2
Arabian Light		5.3		8.5	31.3	7.6
Venezualan						
Tia Juana		2.1				31.7
Algerian						
Hassi Messaoud						21.5
Mixed Canadian			17.8	10.3	7.1	
Indonesian						
Minas					10.4	
Subtotal foreign crude	0.0	11.2	17.8	18.8	48.8	77.0
Total crude - %	100.0	100.0	100.0	100.0	100.0	100.0
Sulfur content % weight						
Model average[a]	.331	.765	.367	1.130	1.251	.789
Industry average[b]	.4	.72	.37	1.17	1.30	.73

[a] Based on weighted average of sulfur content of crudes in model runs (Appendix H).

[b] Reference – transmitted to ADL by EPA on 1-22-75.

Table I-8. TEXAS GULF CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity[a], 1974				Unit capacity[a], 1973				Unit capacity[a,b]		Model limit MB/SD
	Exxon Baytown, Texas	Gulf Port Arthur, Texas	Mobil Beaumont, Texas	1974 Average	Exxon Baytown, Texas	Gulf Port Arthur, Texas	Mobil Beaumont, Texas	1973 Average	1973/ 1974 Average	% Crude cap.	
Crude capacity, Bbls/CD	400,000	312,100	325,000	345,700	350,000	312,100	335,000	332,367	339,034		
	420,000	319,000	335,000	358,000	365,000	319,000	350,000	344,667	351,333		
Vacuum dist.	180,000	147,400	103,000	143,467	150,000	147,400	103,000	133,467	138,467	39.4	
Thermal											
–Visb.							12,000	4,000	2,000	0.6	
–Delayed coke		30,000	33,000	21,000		30,000	33,000	21,000	21,000	6.0	
Catalytic cracking	124,000	120,000	95,000	113,000	135,000	120,000	95,000	116,667	114,834	32.7	114.8
Catalytic reforming	88,000	65,000	94,000	82,333	88,000	65,000	94,000	82,333	82,333	23.4	82.3
Hydrocracking											
–Dist.	20,000	15,000	29,000	21,333	20,000	15,000	29,000	21,333	21,333	6.1	21.3
–Residual											
Hydrofining											
–Hvy gas oil	48,000			16,000	53,000			17,667	16,834	4.8	
–Resid, visb.											
–Cat feed & cycle											
–Distillate		65,000		21,667		65,000		21,667	21,667	6.2	
Hydrotreat											
–Reform feed	90,000	65,000	83 000	79,333	90,000	65,000	83,000	79,333	79,333	22.6	
–Naphtha	15,000			5,000	32,000			10,667	7,833	2.2	
–Olef/arom sat		1,200		400		1,200		400	400	0.1	
–S.R. distill.											38.5
–Lubes	41,000	13,900		18,300	39,500	13,900		17,800	18,050	5.1	18.1
–Other dist.	109,000		42,000	50,333	34,000		42,000	42,000	46,167	13.1	46.2
–Other	8,500[c]			2,833	8,500			2,833	2,833	0.8	
Alkylation	26,000	20,000	16,500	20,833	26,000	20,000	16,500	20,833	20,833	5.9	20.8
Arom/isom											
–BTX		2,700		900		2,700		900	900	0.3	
–HDA											
–Cyclohex		2,500		833		2,500		833	833	0.2	
–C4 Feed											
–C5 Feed		7,200		2,400		7,200		2,400	2,400	0.7	2.4
–C5/C6 Feed											
Lubes	25,000	13,200	8,800	15,667	25,000	13,200	8,800	15,667	15,667	4.4	
Asphalt	12,000		100	4,033	12,000		100	4,033	4,033	1.1	
Coke–tons/day		1,390	1,200	863		1,390	1,200	863	863	–	

[a] Bbls/SD unless otherwise noted.

[b] Used for cluster model.

[c] Solvents.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table I-9. LOUISIANA GULF CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity,[a] 1974				Unit capacity,[a] 1973				Unit capacity[a,b]		Model limit MB/SD
	Gulf Alliance, La.	Shell Norco, La.	Citgo L. Chas., La.	1974 Average	Gulf Alliance, La.	Shell Norco, La.	Citgo L. Chas., La.	1973 Average	1973/ 1974 Average	% Crude cap.	
Crude capacity, Bbls/CD	180,400	240,000	268,000	229,467	174,000	240,000	240,000	218,000	223,734		
	186,000	250,000	N.R.	–	180,000	250,000	245,000	225,000	–		
Vacuum dist.	55,000	90,000	60,000	68,333	54,000	90,000	78,000	74,000	71,167	30.8	
Thermal											
–Visb.											
–Delayed coke	16,000	18,000	28,000	20,667	16,000	17,000	25,000	19,333	20,000	8.7	
Catalytic cracking	78,000	95,000	125,000	99,333	75,000	85,000	112,500	90,833	95,000	41.1	95.0
Catalytic reforming	37,500	41,500	46,000	41,667	37,500	43,000	39,000	39,833	40,750	17.6	40.5
Hydrocracking											
–Dist.		28,000		9,333		29,400		9,800	9,566	4.1	9.5
–Residual							6,000	2,000	1,000	0.4	
Hydrofining											
–Residual			6,000	2,000					1,000	0.4	
–Hvy gas oil											
–Resid. visb.											
–Cat feed & cycle	16,000	25,000	30,000	23,667	16,000	25,000		13,667	18,667	8.1	29.0
–Distillate	22,000			7,333	22,000			7,333	7,333	3.2	7.3
Hydrotreat											
–Reform feed	41,000		46,000	29,000	41,000		16,300	19,100	24,050	10.4	
–Naphtha		26,000		8,667		26,000	11,200	12,400	10,534	4.6	
–Olef/arom sat											
–S.R. distill.			14,000	4,667					2,333	1.0	
Lubes											
–Other dist.											
–Other							24,000	8,000	4,000	1.7	
Alkylation	28,400	14,100	35,300	25,933	28,400	14,100	26,000	22,833	24,383	10.6	24.0
Arom/isom											
–BTX	11,100			3,700	11,100			3,700	3,700	1.6	
–HDA	5,400			1,800	5,400			1,800	1,800	0.8	
–Cyclohex											
–C4 Feed											
–C5 Feed											
–C5/C6 Feed											
Lubes			7,000	2,333			10,000	3,333	2,833	1.2	
Asphalt		6,000		2,000		6,000		2,000	2,000	0.9	
Coke–tons/day	840	900	1,000	913	840	900	895	878	896	–	

[a] Bbls/SD unless otherwise noted.

[b] Used in cluster model.

Reference: *Oil and Gas Journal,* April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table I-10. LARGE MIDWEST CLUSTER PROCESS CONFIGURATION

Unit type	Unit capacity[a], 1974				Unit capacity[a], 1973				Unit capacity[a,b]		Model limit MB/SD
	Mobil Joliet, Ill.	Union Lemont, Ill.	Arco E. Chic., Ill.	1974 Average	Mobil Joliet, Ill.	Union Lemont, Ill.	Arco E. Chic., Ill.	1973 Average	1973/ 1974 Average	% Crude cap.	
Crude capacity, Bbls/CD	175,000	152,000	126,000	151,000	160,000	140,000	135,000	145,000	148,000		
	186,000	N.R.	140,000		164,000	N.R.	140,000				
Vacuum dist.	82,000	55,000	70,000	69,000	72,500	55,000	70,000	65,833	67,417	43.4	
Thermal											
–Gas oil						19,000		6,333	3,167	2.0	
–Visb.											
–Delayed coke	28,000	19,500		15,833	28,000			9,333	12,583	10.1	
Catalytic cracking	66,000	52,000	48,000	55,333	66,000	50,000	48,000	54,667	55,000	35.4	55.0
Catalytic reforming	46,200	32,000	20,000	32,733	46,200	32,000	20,000	32,733	32,733	21.1	32.7
Hydrofining											
–Hvy gas oil											
–Resid. visb.											
–Cat feed & cycle											
–Distillate	69,000		25,000	31,333			25,000	8,333	19,833	12.7	
Hydrotreat											
–Reform feed	67,000	32,000	20,000	39,667	53,000	32,000	20,000	35,000	37,334	24.0	
–Naphtha		2,700	2,000	1,567		2,000	2,000	1,333	1,450	.9	
–Olef/arom sat		4,500		1,500		5,300		1,767	1,634	1.1	
–S.R. distill.		7,000		2,333		7,000		2,333	2,333	1.5	22.7
–Lubes											
–Other dist.		34,500		11,500	54,000			18,000	14,750	12.9	20.0
–Other		2,500[c]		833		37,000		12,333	6,583	4.2	
Alkylation	22,000	12,800	6,000	13,600	18,000	16,000	6,000	13,333	13,467	8.6	13.5
Arom/isom											
–BTX		3,300		1,100		3,200		1,067	1,084	.7	
–HDA											
–Cyclohex											
–C4 Feed											
–C5 Feed											
–C5/C6 Feed											
Lubes											
Asphalt		2,000	10,400	4,133			10,400	3,467	3,800	2.4	
Coke–tons/day	1,700	1,000		900	1,700	1,000		900	900	–	

[a] Bbls/SD unless otherwise noted.

[b] Used for cluster model.

[c] Benzene concentrate.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table I-11. SMALL MIDCONTINENT CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity[a], 1974				Unit capacity[a], 1973				Unit capacity[a,b]		Model limit MB/SD
	Skelly El Dorado, Kan.	Gulf Toledo, Ohio	Champlin Enid, Okla.	1974 Average	Skelly El Dorado, Kan	Gulf Toledo, Ohio	Champlin Enid, Okla.	1973 Average	1973/ 1974 Average	% Crude cap.	
Crude capacity, Bbls/CD	73,700	50,300	49,500	57,833	67,000	48,800	48,000	54,600	56,217		
Vacuum dist.	75,000	51,000	52,000	59,333	70,000	50,000	50,000	56,667	58,000		
Thermal	23,000	12,500	18,000	17,833	23,000	12,300	24,000	19,767	18,800	32.4	
–Visb.											
–Delayed coke	9,800		3,700	4,500	9,800		4,000	4,600	4,550	7.8	
Catalytic cracking	30,000	20,000	19,500	23,167	30,000	18,500	19,000	22,500	22,834	39.4	22.8
Catalytic reforming	21,500	11,000	15,000	15,833	20,000	10,500	15,000	15,167	15,500	26.7	15.5
Hydrofining											
–Hvy gas oil											
–Resid. visb.											
–Cat feed & cycle											
–Distillate		5,000		1,667		5,000		1,667	1,667	2.9	
Hydrotreat											
–Reform feed	23,000	11,000	20,400	18,133	23,000	10,500	15,000	16,167	17,150	29.6	
–Naphtha											
–Olef/arom sat	4,300			1,433	4,300			1,433	1,433	2.5	
–S.R. distill.											1.7
–Lubes											
–Other dist.											
–Other							5,000[c]	1,667	834	1.4	
Alkylation	6,000	5,500	4,500	5,333	6,000	5,100	4,400	5,167	5,250	9.1	5.3
Arom/isom											
–BTX	1,400			467	1,400			467	467	0.8	
–HDA											
–Cyclohex											
–C4 Feed											
–C5 Feed											
–C5/C6 Feed			6,000	2,000			5,000	1,667	1,834	3.2	1.8
Lubes			1,100	367			1,200	400	384	0.7	
Asphalt		2,000	1,400	1,133	3,000	2,000	2,000	2,333	1,733	3.0	
Coke—tons/day	500		165	222	500		158	219	221	–	

[a] Bbls/SD unless otherwise noted.

[b] Used for cluster model.

[c] Isom feed.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table I-12. WEST COAST CLUSTER MODEL PROCESSING CONFIGURATION

Unit type	Unit capacity[a], 1974				Unit capacity[a], 1973				Unit capacity[a,b]		Model limit MB/SD
	Mobil Torrance, Calif.	Arco Carson, Calif.	Socal El Segundo, Calif.	1974 Average	Mobil Torrance, Calif.	Arco Carson, Calif.	Socal El Segundo, Calif.	1973 Average	1973/1974 Average	% Crude cap.	
Crude capacity, Bbls/CD	123,500	165,000	230,000	172,833	123,500	165,000	N.R.				
	130,000	173,000	N.R.		130,000	173,000	220,000	174,333			
Vacuum dist.	95,000	93,000	103,000	97,000	95,000	93,000	103,000	97,000	97,000	54.5	
Thermal											
–Gas oil		12,500		4,167		23,000		7,667	5,917	3.3	
–Visb.	16,000	42,000		19,333	16,000	37,000		17,667	18,500	10.4	
–Delayed coke	46,640	30,000	54,000	43,547	46,640	25,500	50,000	40,713	42,130	23.7	
Catalytic cracking	56,000	57,000	43,500	52,167	56,000	57,000	40,000	51,000	51,584	29.0	51.6
Catalytic reforming	36,000	34,000	60,000	43,333	36,000	32,000	62,000	43,333	43,333	24.3	43.3
Hydrocracking											
–Dist.	18,000	19,700	49,000	28,900	18,000	17,000	45,000	26,667	27,784	15.6	27.8
–Residual											
Hydrofining											
–Hvy gas oil											
–Resid. visb.											
–Cat feed & cycle											
–Distillate		18,000		6,000		18,000		6,000	6,000	3.4	
Hydrotreat											
–Reform feed	23,000	34,000	40,000	32,333	23,000	32,000	40,000	31,667	32,000	18.0	
–Naphtha		18,000		6,000					3,000	1.7	
–Olef/arom sat	15,000			5,000	15,000	18,000		11,000	8,000	4.5	
–S.R. distill.			12,000	4,000			12,000	4,000	4,000	2.2	27.0
–Lubes											
–Other dist.	25,000			8,333	23,000			7,667	8,000	4.5	8.0
–Other			18,000[c]	6,000			18,000	6,000	6,000	3.4	
Alkylation	10,500	7,200	5,900	7,867	10,500	7,200	5,400	7,700	7,784	4.4	7.8
Arom/isom											
–BTX		2,490	—	830	—	2,490	1,500	1,330	1,080	0.7	
–HDA											
–Cyclohex											
–C4 Feed			1,500	500					250	0.1	
–C5 Feed											
–C6/C5 Feed											
Lubes											
Asphalt			8,300	2,767			8,300	2,767	2,767	1.6	
Coke–tons/day	2,800	1,800	2,200	2,267	2,800	1,650	2,200	2,217	2,242	–	

[a] Bbls/SD unless otherwise noted.

[b] Used for cluster model

[c] Jet fuel.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

Table I-13. EAST COAST CLUSTER PROCESSING CONFIGURATION

Unit type	Unit capacity[a], 1974				Unit capacity[a], 1973				Unit capacity[a,b]		Model limit MB/SD
	Arco Phil., Pa.	Sun Marcus Hook, Pa.	Exxon Linden, N.J.	1974 Average	Arco Phil., Pa.	Sun Marcus Hook, Pa.	Exxon Linden, N.J.	1973 Average	1973/ 1974 Average	% Crude cap.	
Crude capacity, Bbls/CD	186,000	165,000	275,000	208,333	160,000	163,000	255,000	192,667	200,500		
	195,000	180,000	286,000	220,333	165,000	180,000	268,000	204,333	212,333		
Vacuum dist.	57,000	48,000	143,000	82,667	83,000	48,000	140,000	90,333	86,500	40.7	
Thermal											
–Visb.											
–Delayed coke											
Catalytic cracking		75,000	120,000	65,000	36,000	75,000	125,000	78,667	71,834	33.8	71.8
Catalytic reforming	60,000	45,000	42,000	49,000	60,000	43,000	46,000	49,667	49,334	23.2	49.3
Hydrocracking											
–Dist.	30,000	—		10,000	30,000	—		10,000	10,000	4.7	10.0
–Other											
Hydrofining											
–Hvy gas oil	32,000		50,000	27,333			50,000	16,667	22,000	10.4	
–Resid. visb.											
–Cat feed & cycle											
–Distillate	41,000			13,667	34,000			11,333	12,500	5.9	
Hydrotreat											
–Reform feed	54,000	35,000	42,000	43,667	53,000	35,000	46,000	44,667	44,167	20.8	
–Naphtha			14,000	4,667			14,000	4,667	4,667	2.2	
–Olef/arom sat											
–S.R. distill.											34.5
–Lubes		10,000		3,333		10,000		3,333	3,333	1.6	
–Other dist.			39,000	13,000			37,000	12,333	12,667	6.0	17.0
–Other		10,000[c]		3,333		16,000[c]		5,333	4,333	2.0	
Alkylation		12,000	8,500	6,833	7,000	12,000	10,700	9,900	8,367	3.9	8.4
Arom/isom											
–BTX		5,300		1,767		5,300		1,767	1,767	.8	
–HDA											
–Cyclohex											
–C4 Feed											
–C5 Feed											
–C5/C6 Feed											
Lubes		17,000		5,667		17,000		5,667	5,667	2.7	
Asphalt	19,500	12,000	46,000	25,833	17,000	12,000	46,000	25,000	25,416	12.0	
Coke—tons/day											

[a] Bbls/SD unless otherwise noted.

[b] Used for cluster model.

[c] Furnace oil.

Reference: *Oil and Gas Journal*, April 2, 1973.
Oil and Gas Journal, April 1, 1974.

calibration effort, since it was not known if any operating units were, in fact, shut down for maintenance during 1973. When this processing configuration capacity data was used for computer runs for 1977, 1980, and 1985, appropriate stream-day-factors were applied.

A stream day capacity limit for a conversion processing unit such as catalytic cracking or reforming is only meaningful at a given operating severity level. The maximum feed rate that a unit can process will increase as severity declines. The Oil and Gas Journal capacity data was assumed to be applicable at 65% volume conversion for catalytic cracking and 95 RON for catalytic reforming. Feed capacity adjustments for severity were used in the calibration runs.

Since the Oil and Gas Journal does not provide capacity data on hydrogen manufacture or sulfur recovery facilities, no limits on these operations were imposed in the calibration runs.

3. Product Data

As part of this project, EPA obtained from each individual oil company the average gasoline grade distribution for calendar year 1973 and associated octane levels/lead additions for each grade, shown in Table I-14. Also shown in Table I-14 are total gasoline volumes and average sulfur contents as supplied by the individual companies and compiled by the EPA. In some cases, the gasoline volumes deviated from the information received from the BOM - for the Texas Gulf Coast and West Coast, the industry production data was about 3 MB/CD below BOM statistics. For these cases, the BOM data was used in the calibration effort.

Table I-15 provides other key product specifications used in the calibration runs. Distillate fuel sulfur specifications were allowed to vary from model to model in an effort to achieve reasonable utilization of existing desulfurization capacity. Since U. S. refiners did not desulfurize residual stocks in 1973, sulfur specifications were relatively "loose" for this product. After discussions with several oil companies, it was felt that LPG for petrochemicals feedstock use would be met by a fixed blend of 80% mixed C_3/C_4 olefins, 16% propane and 2% each of iso and normal butane.

Table I-14. CLUSTER MODEL GASOLINE PRODUCTION AND PROPERTIES – 1973[a,b]

	Texas Gulf	Louisiana Gulf	East Coast	Small Midcontinent	Large Midwest	West Coast
Gasoline volume						
MBPY	59,360	43,335	38,301	13,663		23,904
MB/CD	162.6	118.8	104.9	37.4	81.7	65.5
Sulfur content, %	.041	.044	.023	.034	.082	.07
Pool octanes						
Clear RON	88.2	88.95	87.8	86.1	87.9	90.4
Leaded RON	95.5					
Clear MON	79.7	81.63	80.0	80.0	79.6	81.5
Leaded MON	88.3					
Lead g/gal	2.09	1.77	1.82	1.58	1.74	1.94
Grade distribution and octane						
Premium, %	28.46	34.54	28.8	20.67	20.2	53.57
Premium, octane						
Leaded RON	99.5	99.8	100.5	98.9	99.2	99.3
Clear RON	92.9			90.4		
Leaded MON	92.0	92.2	92.5	94.0	92.6	90.2
Clear MON	83.2			80.0		
Lead g/gal	2.43	1.99	2.27	2.30	1.99	2.6
Regular, %	60.68	58.05	64.9	75.70	79.6	45.07
Regular, octane						
Leaded RON	93.8	93.6	94.1	92.2	94.0	93.4
Clear RON	85.5			84.7		
Leaded MON	86.8	86.6	86.1	86.1	86.2	84.5
Clear MON	77.3			78.6		
Lead g/gal	2.22	1.81	1.76	1.44	1.68	1.21
Low lead, %	10.86	7.42	6.3	3.63	0.2	0.78
Low lead, octane						
Leaded RON	93.7	94.3	96.6	91.0	94.6	94.0
Clear RON	91.7			88.7		
Leaded MON	86.3	85.7	86.8	86.0	88.0	85.1
Clear MON	83.4			79.6		
Lead g/gal	0.44	0.5	0.4	0.36	0.3	0.3
Unleaded, %	–	–	–	–	–	0.58
Unleaded, octane						
RON	–	–	–	–	–	92.0
MON	–	–	–	–	–	82.7

[a] Reference: transmitted to ADL by EPA on 1-22-75

[b] No entry indicates data not reported.

Table 1-15. KEY PRODUCT SPECIFICATIONS

Product	Specifications	All clusters	Texas[a] Gulf	La. Gulf	Large Midwest	Small Midcontinent	East Coast	West Coast
Motor gasoline	Maximum vapor pressure (RVP)	10.5						
	Maximum lead addition (g/gal)	3.17						
Kerosene	Maximum sulfur (% Wt)	0.1						
	Maximum gravity (°API)	48.0						
Jet fuel	Maximum sulfur (% Wt)	0.1						
	Maximum gravity (°API)	46.0						
	Minimum smoke point (mm)	20.0						
Residual fuel oil	Maximum viscosity (Refutas @ 122°F)	38.0						
	Maximum sulfur (% Wt)		.78	.75	.78	.78	.78	1.90
Distillate fuel oil	Maximum sulfur (% Wt)		.17	.17	.30	.30	.10	.17

[a] Volatility specifications for this cluster used as shown in Table C-1.

4. Calibration Economic Data

As was stated previously, for the calibration runs each individual product demand was fixed as well as non-crude inputs. For each cluster model, one important crude, a reference crude, was allowed to vary to balance product demands and internal fuel requirements. Other crude inputs were fixed.

The reference crudes used were:

Cluster Model	Reference Crude
Louisiana Gulf	Louisiana
Texas Gulf	West Texas Sour
Large Midwest	West Texas Sour
Small Midcontinent	Oklahoma
East Coast	Tia Juana
West Coast	Wilmington

The reference crude was assumed to cost $4.00/bbl. for all models except the East Coast, which used $4.25/bbl.

Maximum natural gas availability was established for each cluster model based on the BOM data presented in Table I-3 under category A-4. In some cases, this value was increased somewhat to account for natural gas consumed in hydrogen manufacture. The model could purchase up to the specified maximum at the prices indicated as follows:

Cluster Model	Maximum Natural Gas Availability MB/CD (FOE)	Price $ per bbl. (FOE)
Louisiana	5.4	1.89
Texas Gulf	29.24	1.89
Large Midwest	.24	2.52
Small Midcontinent	2.046	1.89
East Coast	2.50	3.15
West Coast	6.39	1.89

Full refinery operating costs were used for the optimization of the calibration runs. Unit costs for all models included purchased electricity at 1.2¢ per KWH, cooling water at 4¢ per 1,000 gal., and $430.00 per daily shift position for operating labor. Capital charge (annualized capital recovery rate) for the existing capacity in the cluster model was reduced to 1/10th the normal level (25%) which would be assessed to new plant construction. (This resulted in a 2.5% per year annual charge of the plant capital value in 1973 dollars.)

In theory, the day-to-day optimization and operation of any manufacturing facilities should consider the capital charges associated with existing plant investments as "sunk" capital, and thus not be a factor for influencing operating decisions. However, there can be tax savings associated with early tax write-offs of existing facilities and the land area occupied by refinery processing units undoubtedly has value. Thus, we feel a relatively nominal 10% of full capital charge to be a reasonable approach in the optimization of existing facilities.

B. CALIBRATION RESULTS FOR CLUSTER MODELS

There are four main areas in which one can compare the degree of calibration for the cluster models. These are:

- Overall Refinery Material Balance (i.e., volume of the crude intake required to balance specified product demands and internal fuel requirements)
- Refinery Energy Consumption
- Processing Configuration, Throughputs and Operating Severities
- Key Product Properties (i.e., gasoline clear pool octanes, lead levels, etc.)

As noted previously in this Appendix, the EPA obtained specific operating data from each individual refinery within each cluster model, and EPA averaged the industry statistics which were then transmitted to ADL. Table I-16 presents this data as received from the EPA.

Refinery flow diagrams resulting from the calibration runs are shown in Figures I-1 through I-6.

Table I-16. CLUSTER MODEL PROCESSING DATA[a] – 1973

	Texas Gulf			Louisiana Gulf			Small Midcontinent			Large Midwest			East Coast			West Coast		
Unit	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation
Crude distillation																		
Atmospheric																		
MB/CD	331.4			216.8			56.6			145.5			186.5			162.0		
MB/SD	351.3			232.2			57.8			156.3			207.3			177.6		
		94.3			93.4			97.8			93.1			90.0			91.2	
Vacuum																		
MB/CD	129.2			66.1			18.5			58.3			76.1			84.1		
MB/SD	138.4			71.2			19.2			67.4			90.8			97.0		
		93.3			92.9			96.4			86.4			86.0			86.7	
Thermal Operations																		
Delayed coker																		
MB/CD	19.3			19.1			4.6			13.6						38.4		
MB/SD	20.5			20.5			4.6			15.8						42.1		
		94.2			93.0			100.0			85.9						91.2	
Catalytic Cracker																		
MB/CD	100.4			90.0			20.2			51.2			73.1			44.5		
MB/SD	114.8			96.1			23.0			56.5			76.5			51.6		
		87.5			94.6			88.1			92.3			95.6			86.2	
Conversion			73.6			70.0			76.5			74.9			71.6			61.6
Catalytic reformer																		
MB/CD	73.8			30.2			11.7			27.8			36.3			34.7		
MB/SD	82.3			36.9			13.8			32.7			43.4			43.0		
		89.6			81.7			85.0			84.8			83.7			80.7	
Severity, RON clear			94.8			92.3			85.8			90.7			93.4			95.9
Hydrocracker																		
MB/CD	14.7			6.3									7.4			23.2		
MB/SD	21.3			9.6									10.0			27.8		
		69.0			66.2									74.0			83.4	
Alkylation																		
MB/CD	17.8			17.7			5.0			11.4			6.9			5.6		
MB/SD	20.8			21.1			5.3			12.9			9.2			7.8		
		85.6			83.8			95.6			87.9			75.0			71.5	

Table I-16 (continued). CLUSTER MODEL PROCESSING DATA[a] – 1973

Unit	Texas Gulf			Louisiana Gulf			Small Midcontinent			Large Midwest			East Coast			West Coast		
	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation	Unit Capacity	% Utilized	Operation
Aromatics																		
C_5 isomerization																		
MB/CD	2.36						1.4											
MB/SD	2.40						1.8											
		98.6						76.6										
Benzene (HDA)																		
MB/CD	0.83			1.5														
MB/SD	0.90			1.8														
		92.5			85.2													
BTX reformer																		
MB/CD				2.7			1.6						4.4					
MB/SD				3.7			1.8						6.1					
					73.0			88.9						73.0				
UDEX																		
MB/CD							0.30						4.5					
MB/SD							0.47						5.0					
								64.9						90.0				

Reference: transmitted to ADL by EPA on 1-25-75.

a. MB/CD data supplied by industry to EPA.

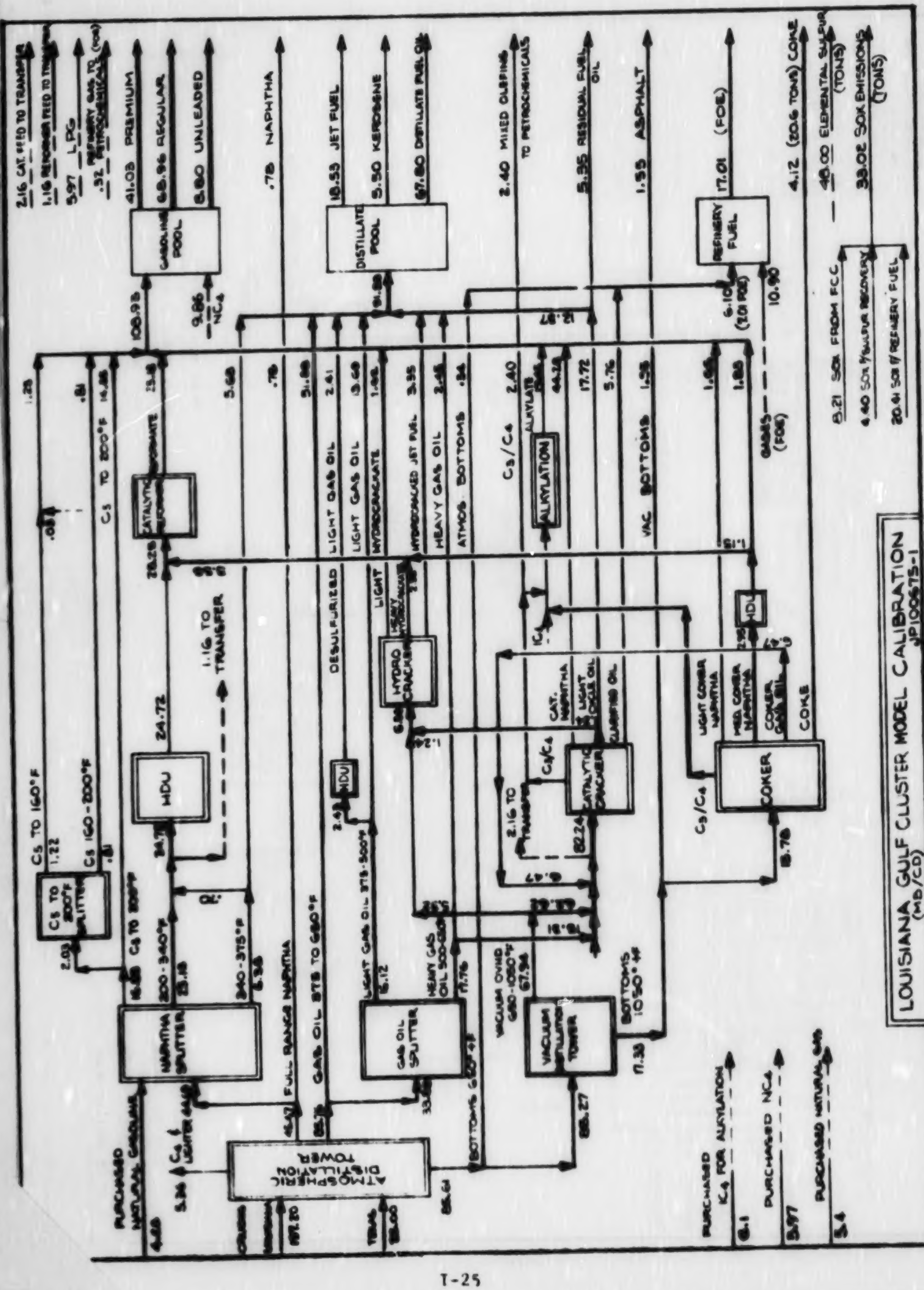

LOUISIANA GULF CLUSTER MODEL CALIBRATION
(MB/CD)
JP100575-1
ATMOSPHERIC DISTILLATION TOWER
NAPHTHA SPLITTER
GAS OIL SPLITTER
VACUUM DISTILLATION TOWER
C5 TO 200°F SPLITTER
MDU
HDU
CATALYTIC REFORMER
HYDRO CRACKER
CATALYTIC CRACKER
ALKYLATION
COKER
GASOLINE POOL
DISTILLATE POOL
REFINERY FUEL
PURCHASED NATURAL GASOLINE
CRUDE
TEXAS
C4 & LIGHTER GAS
PURCHASED iC4 FOR ALKYLATION
PURCHASED NC4
PURCHASED NATURAL GAS
FULL RANGE NAPHTHA
GAS OIL 375 TO 650°F
BOTTOMS 650°+F
C5 TO 160°F
C6 160-200°F
C6 TO 200°F
200-340°F
340-375°F
1.16 TO TRANSFER
LIGHT GAS OIL 375-500°F
HEAVY GAS OIL 500-650°F
VACUUM OVHD 650-1050°F
BOTTOMS 1050°+F
DESULFURIZED
LIGHT
HEAVY HYDROCRACKATE
2.16 TO TRANSFER
C3/C4
iC4
CAT. NAPHTHA
LIGHT CYCLE OIL
CLARIFIED OIL
LIGHT COKER NAPHTHA
MED. COKER NAPHTHA
COKER GAS OIL
COKE
C5 TO 200°F
REFORMATE
LIGHT GAS OIL
HYDROCRACKATE
HYDROCRACKED JET FUEL
HEAVY GAS OIL
ATMOS. BOTTOMS
ALKYLATE
VAC BOTTOMS
GASES (FOE)
NC4
2.16 CAT. FEED TO TRANSFER
1.16 REFORMER FEED TO TRANSFER
5.97 LPG
.32 REFINERY GAS TO PETROCHEMICALS
41.03 PREMIUM
65.96 REGULAR
8.80 UNLEADED
.78 NAPHTHA
18.53 JET FUEL
5.50 KEROSENE
67.80 DISTILLATE FUEL OIL
2.40 MIXED OLEFINS TO PETROCHEMICALS
5.35 RESIDUAL FUEL OIL
1.55 ASPHALT
17.01 (FOE)
4.12 (20.6 TONS) COKE
48.00 ELEMENTAL SULFUR (TONS)
33.02 SOX EMISSIONS (TONS)
8.21 SOX FROM FCC
4.40 SOX F/SULFUR RECOVERY
20.41 SOX F/REFINERY FUEL
108.93
9.86
91.83
6.10 (9.01 FOE)
10.90
4.16
5.36
197.20
55.00
45.47
85.78
85.64
33.62
85.27
67.94
17.30
16.88
25.18
5.38
2.03
1.22
.81
24.72
17.76
6.12
2.42
52.24
15.78
1.25
16.28
25.8
5.68
31.85
2.41
3.68
1.92
3.35
2.48
.84
2.40
44.28
17.72
5.76
1.35
1.95
1.88
28.25
TJ080-01-02-09

Figure I-1

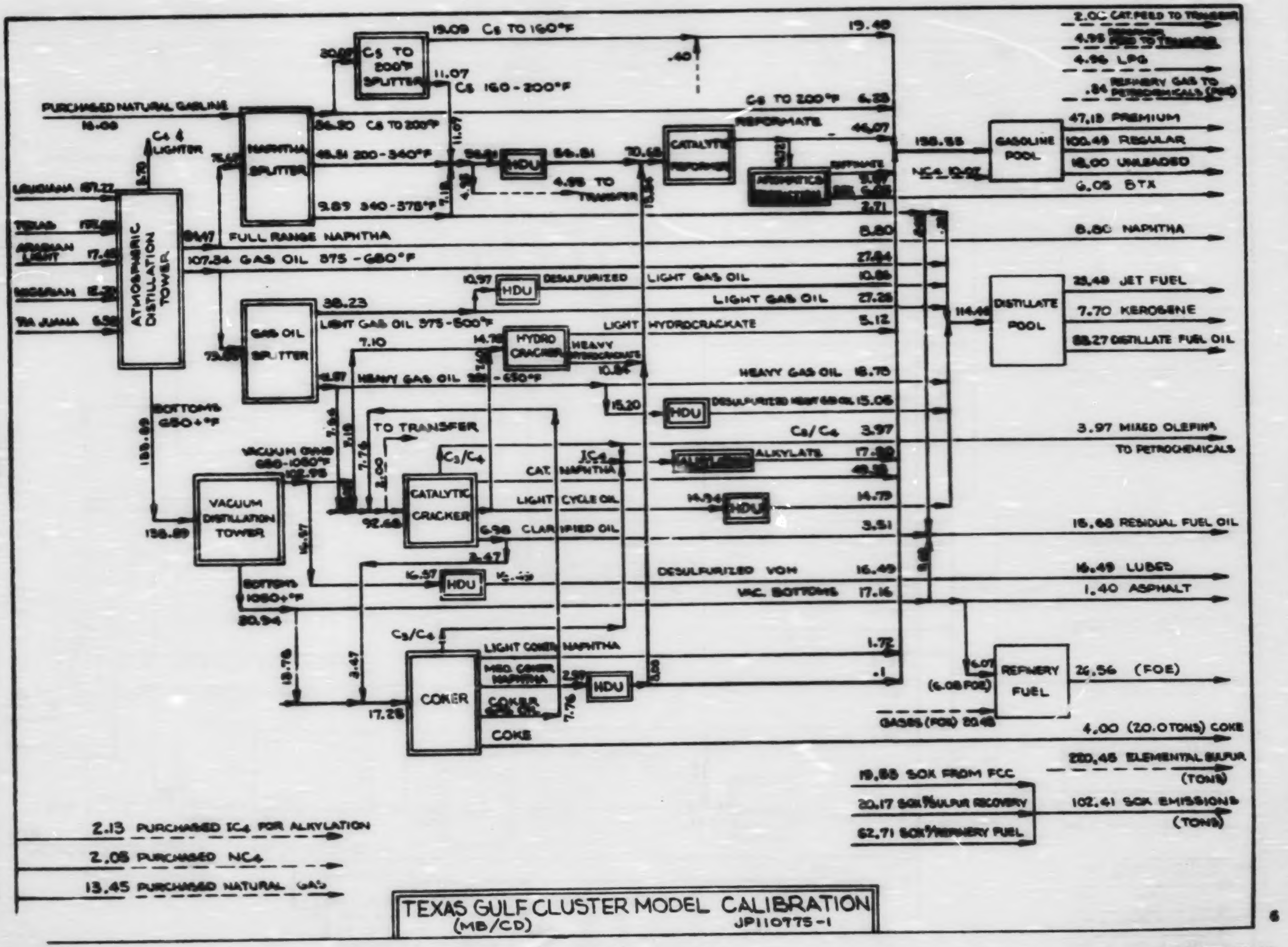
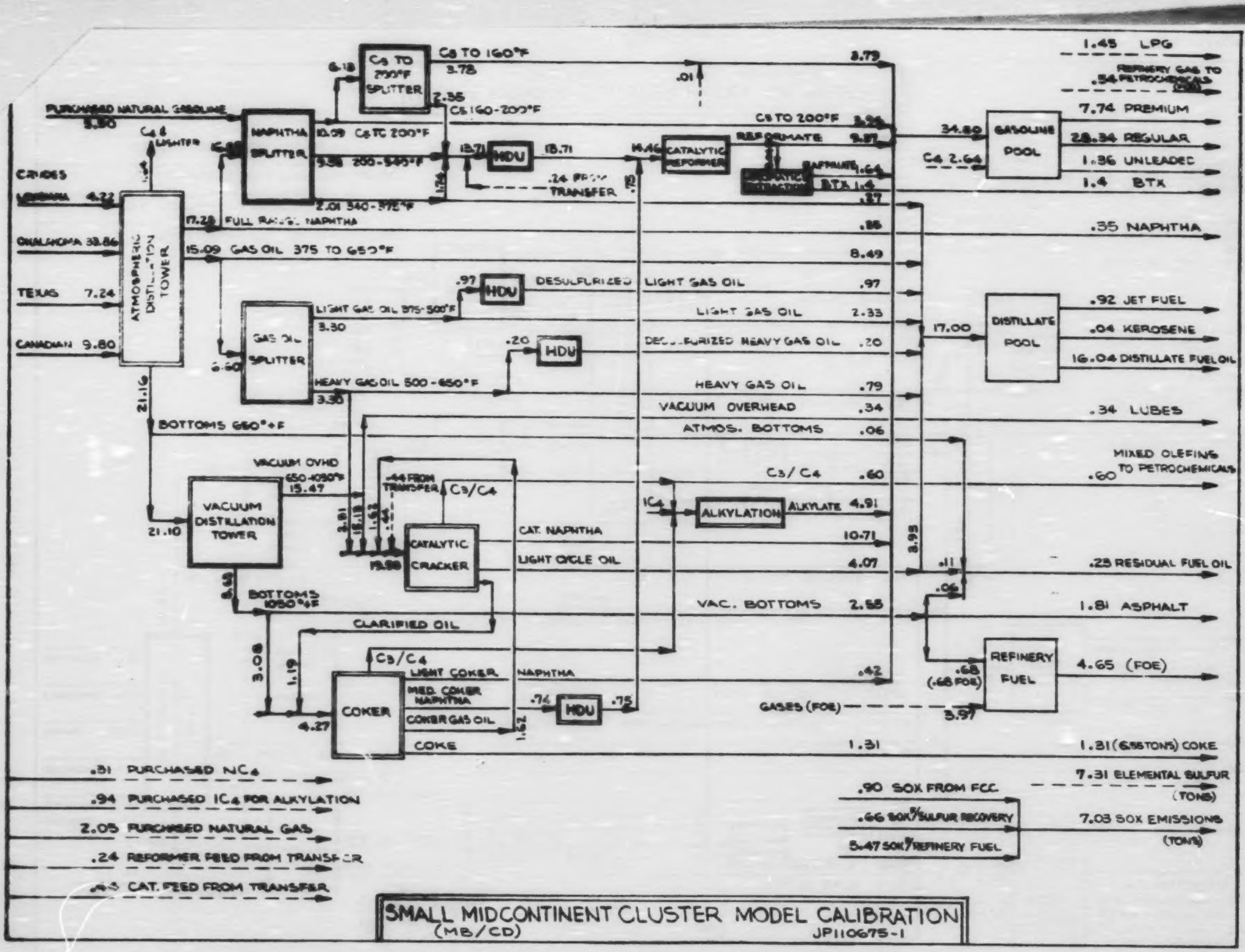

19.09 C5 TO 160°F
19.40
.40
C5 TO 200°F SPLITTER
30.07
11.07
C5 160-200°F
PURCHASED NATURAL GASLINE
16.09
C4 & LIGHTER
8.70
NAPHTHA SPLITTER
75.41
36.30 C6 TO 200°F
11.07
48.81 200-340°F
9.89 340-375°F
7.18
4.95
58.81
HDU
56.81
4.95 TO TRANSFER
70.68
15.86
CATALYTIC REFORMER
C6 TO 200°F 6.23
REFORMATE 46.07
AROMATICS EXTRACTION
RAFFINATE
2.71
LOUISIANA 167.22
TEXAS
ARABIAN LIGHT 17.45
NIGERIAN
TIA JUANA
ATMOSPHERIC DISTILLATION TOWER
FULL RANGE NAPHTHA
8.80
107.84 GAS OIL 375-680°F
27.84
10.97
HDU
DESULFURIZED LIGHT GAS OIL 10.86
38.23
LIGHT GAS OIL 27.26
GAS OIL SPLITTER
73.81
LIGHT GAS OIL 375-500°F
7.10
14.76
HYDRO CRACKER
HEAVY HYDROCRACKATE
10.84
LIGHT HYDROCRACKATE 5.12
HEAVY GAS OIL -650°F
HEAVY GAS OIL 18.75
15.20
HDU
DESULFURIZED HEAVY GAS OIL 15.06
BOTTOMS 650+°F
188.89
TO TRANSFER
7.94
7.19
7.76
2.00
C3/C4
C3/C4 3.97
IC4
CAT. NAPHTHA
ALKYLATION
ALKYLATE 17.80
VACUUM GAS OIL 650-1050°F
102.95
CATALYTIC CRACKER
92.68
LIGHT CYCLE OIL
14.94
HDU
14.79
6.98 CLARIFIED OIL
3.51
3.47
VACUUM DISTILLATION TOWER
138.89
16.97
16.57
HDU
16.49
DESULFURIZED VOH 16.49
BOTTOMS 1050+°F
20.94
VAC. BOTTOMS 17.16
C3/C4
18.76
3.47
17.28
COKER
LIGHT COKER NAPHTHA 1.72
MED. COKER NAPHTHA
2.97
HDU
3.86
.1
COKER GAS OIL
7.76
COKE
138.55
GASOLINE POOL
NC4 10.07
114.46
DISTILLATE POOL
2.00 CAT.FEED TO TRANSFER
4.95 REFORMER FEED TO TRANSFER
4.96 LPG
.84 REFINERY GAS TO PETROCHEMICALS (FOE)
47.18 PREMIUM
100.49 REGULAR
18.00 UNLEADED
6.05 BTX
8.80 NAPHTHA
23.49 JET FUEL
7.70 KEROSENE
83.27 DISTILLATE FUEL OIL
3.97 MIXED OLEFINS TO PETROCHEMICALS
18.68 RESIDUAL FUEL OIL
16.49 LUBES
1.40 ASPHALT
6.07
(6.08 FOE)
REFINERY FUEL
26.56 (FOE)
GASES (FOE) 20.48
4.00 (20.0 TONS) COKE
19.88 SOX FROM FCC
20.17 SOX/SULFUR RECOVERY
62.71 SOX/REFINERY FUEL
220.48 ELEMENTAL SULFUR (TONS)
102.41 SOX EMISSIONS (TONS)
2.13 PURCHASED IC4 FOR ALKYLATION
2.05 PURCHASED NC4
13.45 PURCHASED NATURAL GAS
TEXAS GULF CLUSTER MODEL CALIBRATION
(MB/CD)
JP110775-1
6

Figure I-2

76086-01,08,09



CRUDES
LOUISIANA 4.22
OKLAHOMA 33.86
TEXAS 7.24
CANADIAN 9.80
PURCHASED NATURAL GASOLINE 3.80
C4 & LIGHTER
ATMOSPHERIC DISTILLATION TOWER
17.28 FULL RANGE NAPHTHA
15.09 GAS OIL 375 TO 650°F
21.16
BOTTOMS 650°+F
NAPHTHA SPLITTER
10.09 C5 TO 200°F
3.36 200-340°F
2.01 340-375°F
C5 TO 200°F SPLITTER
6.18
C5 TO 160°F 3.78
2.35 C5 160-200°F
13.71 HDU 13.71
.24 FROM TRANSFER
14.46 CATALYTIC REFORMER
REFORMATE
AROMATIC EXTRACTION
RAFFINATE
BTX 1.4
.01
C5 TO 200°F
GAS OIL SPLITTER
6.60
LIGHT GAS OIL 375-500°F 3.30
HEAVY GAS OIL 500-650°F 3.30
.97 HDU DESULFURIZED LIGHT GAS OIL
LIGHT GAS OIL
.20 HDU DESULFURIZED HEAVY GAS OIL
HEAVY GAS OIL
VACUUM OVERHEAD
ATMOS. BOTTOMS
21.10
VACUUM DISTILLATION TOWER
VACUUM OVHD 650-1050°F 15.47
BOTTOMS 1050°+F
3.08
1.19
.44 FROM TRANSFER
C3/C4
CATALYTIC CRACKER
19.98
CAT. NAPHTHA
LIGHT CYCLE OIL
CLARIFIED OIL
C3/C4
IC4
ALKYLATION
ALKYLATE 4.91
VAC. BOTTOMS
COKER
4.27
C3/C4
LIGHT COKER NAPHTHA
MED. COKER NAPHTHA
COKER GAS OIL
COKE
.74 HDU .75
1.62
GASES (FOE)
3.79
3.96
9.87
.64
1.4
.37
.85
8.49
.97
2.33
.20
.79
.34
.06
.60
4.91
10.71
4.07
2.55
.42
1.31
8.95
.11
.06
.68 (.68 FOE)
3.97
34.80
C4 2.64
GASOLINE POOL
17.00
DISTILLATE POOL
REFINERY FUEL
1.45 LPG
.54 REFINERY GAS TO PETROCHEMICALS (FOE)
7.74 PREMIUM
28.34 REGULAR
1.36 UNLEADED
1.4 BTX
.35 NAPHTHA
.92 JET FUEL
.04 KEROSENE
16.04 DISTILLATE FUEL OIL
.34 LUBES
.60 MIXED OLEFINS TO PETROCHEMICALS
.23 RESIDUAL FUEL OIL
1.81 ASPHALT
4.65 (FOE)
1.31 (6.55 TONS) COKE
.31 PURCHASED NC4
.94 PURCHASED IC4 FOR ALKYLATION
2.05 PURCHASED NATURAL GAS
.24 REFORMER FEED FROM TRANSFER
.44 CAT. FEED FROM TRANSFER
.90 SOX FROM FCC
.66 SOX% SULFUR RECOVERY
5.47 SOX% REFINERY FUEL
7.31 ELEMENTAL SULFUR (TONS)
7.03 SOX EMISSIONS (TONS)
SMALL MIDCONTINENT CLUSTER MODEL CALIBRATION (MB/CD)
JP110675-1

Figure I-3

4

76086-01,-08,09

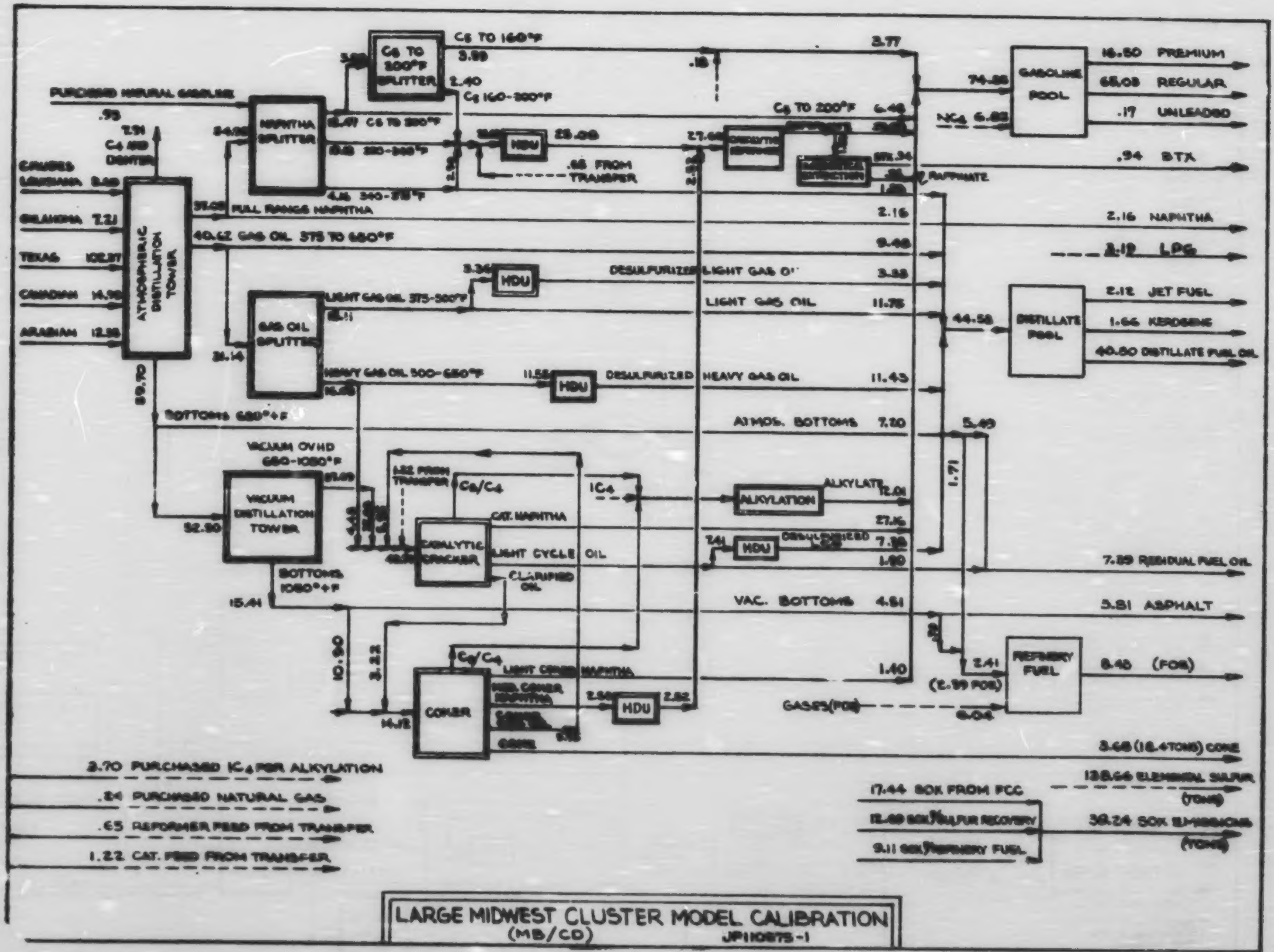

CRUDES
LOUISIANA 8.80
OKLAHOMA 7.21
TEXAS 102.37
CANADIAN 14.98
ARABIAN 12.38
PURCHASED NATURAL GASOLINE
.95
7.31
C4 AND LIGHTER
ATMOSPHERIC DISTILLATION TOWER
37.08 FULL RANGE NAPHTHA
40.62 GAS OIL 375 TO 650°F
89.70
BOTTOMS 650°+F
34.90
NAPHTHA SPLITTER
15.47 C5 TO 200°F
3.89
C6 TO 200°F SPLITTER
C5 TO 160°F
3.89
2.40
C6 160-200°F
340-375°F
4.16
HDU
23.08
.65 FROM TRANSFER
21.14
GAS OIL SPLITTER
LIGHT GAS OIL 375-500°F
3.36
HDU
DESULFURIZED LIGHT GAS OIL
LIGHT GAS OIL
HEAVY GAS OIL 500-650°F
16.65
11.58
HDU
DESULFURIZED HEAVY GAS OIL
VACUUM OVHD 650-1050°F
VACUUM DISTILLATION TOWER
52.90
BOTTOMS 1050°+F
15.41
1.22 FROM TRANSFER
C3/C4
CATALYTIC CRACKER
CAT. NAPHTHA
LIGHT CYCLE OIL
CLARIFIED OIL
IC4
ALKYLATION
ALKYLATE
10.90
3.22
14.12
COKER
C3/C4
LIGHT COKER NAPHTHA
MED. COKER NAPHTHA
COKE
2.50
HDU
2.52
7.41
HDU
DESULFURIZED
27.60
C5 TO 200°F
.18
3.77
6.48
BTX .34
RAFFINATE
2.16
9.48
3.38
11.75
11.43
ATMOS. BOTTOMS 7.20
12.01
27.16
7.38
1.90
VAC. BOTTOMS 4.81
1.40
GASES(FOE)
74.85
NC4 6.82
GASOLINE POOL
16.50 PREMIUM
65.03 REGULAR
.17 UNLEADED
.94 BTX
2.16 NAPHTHA
2.19 LPG
44.58
DISTILLATE POOL
2.12 JET FUEL
1.66 KEROSENE
40.80 DISTILLATE FUEL OIL
5.49
1.71
7.89 RESIDUAL FUEL OIL
5.81 ASPHALT
2.41
(2.39 FOE)
6.04
REFINERY FUEL
8.45 (FOE)
3.68 (18.4 TONS) COKE
2.70 PURCHASED IC4 FOR ALKYLATION
.24 PURCHASED NATURAL GAS
.65 REFORMER FEED FROM TRANSFER
1.22 CAT. FEED FROM TRANSFER
17.44 SOX FROM FCC
12.69 SOX/SULFUR RECOVERY
9.11 SOX/REFINERY FUEL
138.64 ELEMENTAL SULFUR (TONS)
39.24 SOX EMISSIONS (TONS)
LARGE MIDWEST CLUSTER MODEL CALIBRATION
(MB/CD)
JP110875-1

Figure I-4

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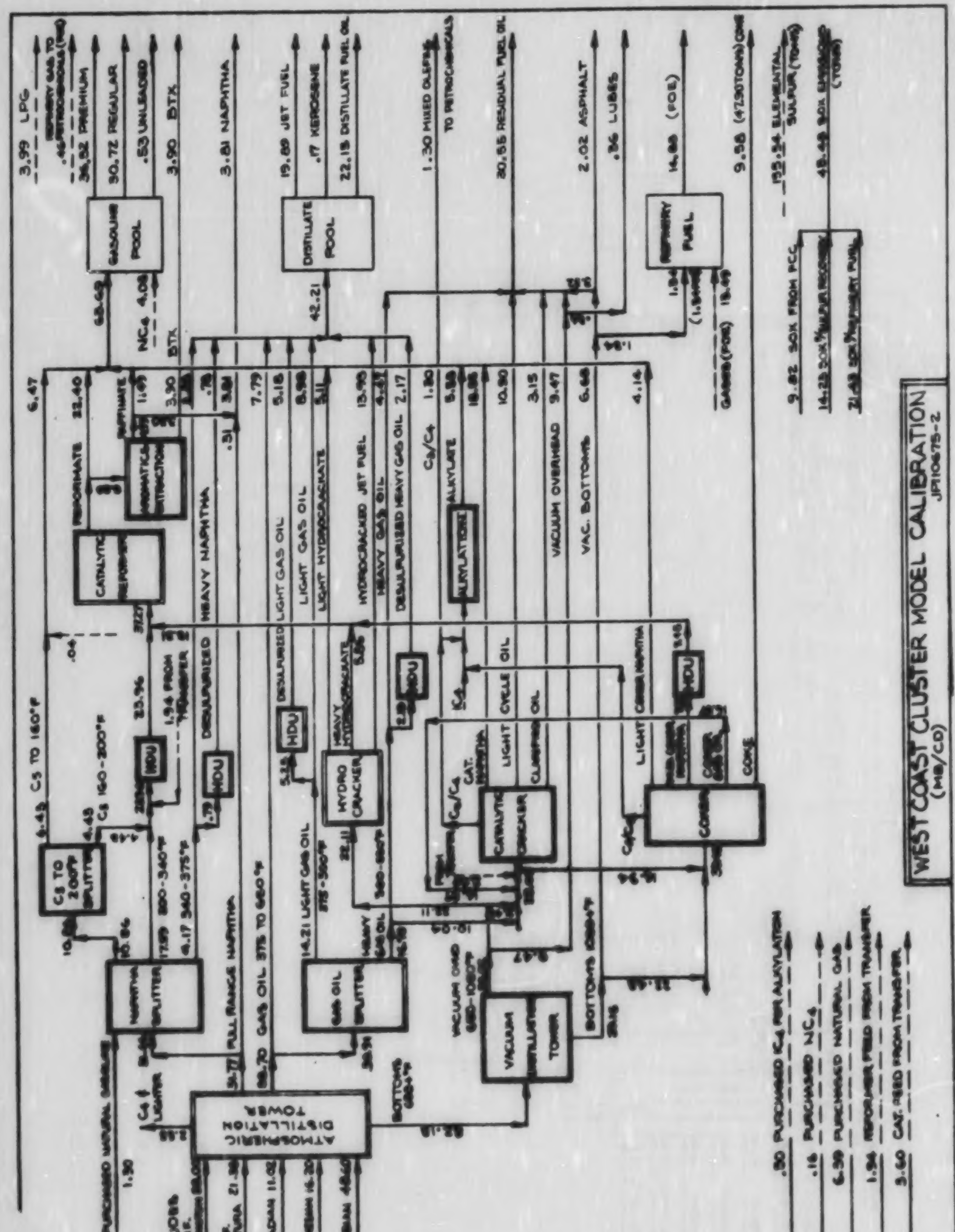
WEST COAST CLUSTER MODEL CALIBRATION
(MB/CD)
JP110675-2
CRUDES
CALIF. VENTURA 21.38
CANADIAN 11.02
INDONESIAN 16.20
ARABIAN 48.60
PURCHASED NATURAL GASOLINE 1.30
C4 & LIGHTER
ATMOSPHERIC DISTILLATION TOWER
NAPHTHA SPLITTER
C5 TO 200°F SPLITTER
C5 TO 160°F
C6 160-200°F
200-340°F
340-375°F
FULL RANGE NAPHTHA
GAS OIL 375 TO 680°F
GAS OIL SPLITTER
LIGHT GAS OIL
HEAVY GAS OIL
BOTTOMS 680+°F
VACUUM DISTILLATION TOWER
VACUUM OVHD 680-1050°F
BOTTOMS 1050+°F
HDU
HYDRO CRACKER
HEAVY HYDROCRACKATE
CATALYTIC REFORMER
REFORMATE
AROMATICS EXTRACTION
RAFFINATE
BTX
CATALYTIC CRACKER
CAT. NAPHTHA
LIGHT CYCLE OIL
CLARIFIED OIL
COKER
LIGHT COKER NAPHTHA
COKER GAS OIL
COKE
ALKYLATION
ALKYLATE
C3/C4
IC4
HEAVY NAPHTHA
DESULFURIZED
LIGHT HYDROCRACKATE
HYDROCRACKED JET FUEL
DESULFURIZED HEAVY GAS OIL
VACUUM OVERHEAD
VAC. BOTTOMS
GASOLINE POOL
DISTILLATE POOL
REFINERY FUEL
NC4
3.99 LPG
.46 REFINERY GAS TO PETROCHEMICALS
36.52 PREMIUM
30.72 REGULAR
.53 UNLEADED
3.90 BTX
3.81 NAPHTHA
19.89 JET FUEL
.17 KEROSENE
22.15 DISTILLATE FUEL OIL
1.30 MIXED OLEFINS TO PETROCHEMICALS
30.55 RESIDUAL FUEL OIL
2.02 ASPHALT
.36 LUBES
14.88 (FOE)
9.58 (4730 TONS) COKE
192.54 ELEMENTAL SULFUR (TONS)
48.49 SOX EMISSIONS (TONS)
9.82 SOX FROM FCC
14.23 SOX SULFUR RECOVERY
21.43 SOX REFINERY FUEL
GASES (FOE) 15.49
.30 PURCHASED IC4 FOR ALKYLATION
.16 PURCHASED NC4
6.39 PURCHASED NATURAL GAS
1.94 REFORMER FEED FROM TRANSFER
3.60 CAT. FEED FROM TRANSFER
TC086-01,08,09

Figure I-5

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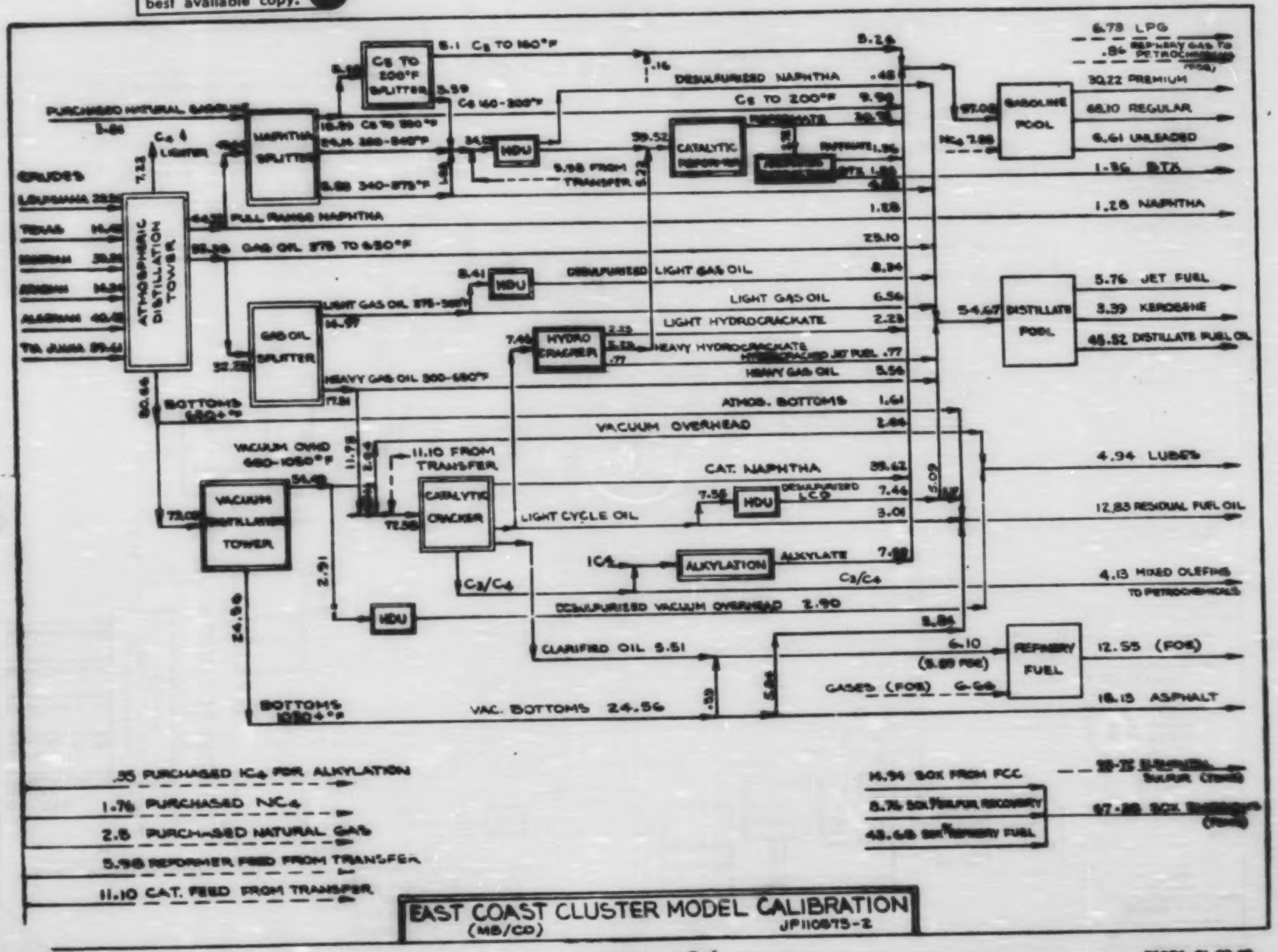
CRUDES
LOUISIANA 22.5
TEXAS 11.4
ALGERIAN 40.8
TIA JUANA 29.61
PURCHASED NATURAL GASOLINE
3.84
ATMOSPHERIC DISTILLATION TOWER
7.23
C4 & LIGHTER
NAPHTHA SPLITTER
16.89 C5 TO 280°F
24.14 280-340°F
3.88 340-375°F
C5 TO 200°F SPLITTER
5.1 C5 TO 160°F
3.59
C5 160-200°F
HDU
34.2
5.98 FROM TRANSFER
.16
DESULFURIZED NAPHTHA
C5 TO 200°F
REFORMATE
CATALYTIC REFORMER
39.52
5.22
RAFFINATE 1.36
BTX
5.26
.48
9.90
1.28
25.10
FULL RANGE NAPHTHA
GAS OIL 375 TO 650°F
8.41
HDU
DESULFURIZED LIGHT GAS OIL
8.34
GAS OIL SPLITTER
32.31
LIGHT GAS OIL 375-500°F
14.97
LIGHT GAS OIL
6.56
LIGHT HYDROCRACKATE
2.23
HYDRO CRACKER
7.46
2.23
5.22
.77
HEAVY HYDROCRACKATE
HYDROCRACKED JET FUEL .77
HEAVY GAS OIL 500-650°F
17.81
HEAVY GAS OIL
5.56
ATMOS. BOTTOMS
1.61
50.66
BOTTOMS 650+°F
VACUUM OVERHEAD
2.86
VACUUM OVHD 650-1050°F
11.79
11.10 FROM TRANSFER
VACUUM DISTILLATION TOWER
73.02
34.46
CATALYTIC CRACKER
72.58
CAT. NAPHTHA
38.62
LIGHT CYCLE OIL
7.56
HDU
DESULFURIZED LCO
7.44
3.01
IC4
ALKYLATION
ALKYLATE
7.88
C3/C4
C3/C4
2.91
HDU
DESULFURIZED VACUUM OVERHEAD 2.90
CLARIFIED OIL 5.51
24.56
BOTTOMS 1050+°F
VAC. BOTTOMS 24.56
.59
5.09
97.08
NC4 7.86
GASOLINE POOL
6.78 LPG
.86 REFINERY GAS TO PETROCHEMICALS (FOE)
30.22 PREMIUM
68.10 REGULAR
6.61 UNLEADED
1.36 BTX
1.28 NAPHTHA
54.67
DISTILLATE POOL
5.76 JET FUEL
3.39 KEROSENE
45.52 DISTILLATE FUEL OIL
4.94 LUBES
12.83 RESIDUAL FUEL OIL
4.13 MIXED OLEFINS TO PETROCHEMICALS
6.10
(5.89 FOE)
GASES (FOE) 6.66
REFINERY FUEL
12.55 (FOE)
18.15 ASPHALT
.55 PURCHASED IC4 FOR ALKYLATION
1.76 PURCHASED NC4
2.8 PURCHASED NATURAL GAS
5.98 REFORMER FEED FROM TRANSFER
11.10 CAT. FEED FROM TRANSFER
14.94 SOX FROM FCC
8.76 SOX FROM SULFUR RECOVERY
43.68 SOX FROM REFINERY FUEL
SULFUR (TONS)
67.38 SOX
(TONS)
78086-01,08,09

EAST COAST CLUSTER MODEL CALIBRATION
(MB/CD)
JP110875-2

Figure I-6

Tables I-17 through I-22 provide key calibration results concerning the first three, of the above four, comparison areas, for each of the cluster models. For example, Table I-18 provides the information for the Texas Gulf cluster. The first element in this table reports the total crude intake for the calendar year 1973 from the BOM data, industry data, and the ADL model run. The BOM crude input data (converted to MB/CD) was obtained from Table I-1 by adding categories 1(a) + 1(b), subtracting category 24 (the ADL model did not manufacture "other" finished products since they could not be discretely identified), plus BOM category 3 (whenever information was reported in this category). The industry data as compiled by the EPA reflects some non-crude material being charged to atmospheric distillation units, but in general, checks reasonably well with BOM statistics.

Next, in Table I-18, is a comparison of refinery energy consumption. The BOM data for purchased natural gas is obtained from Table I-3 assuming 1,000 BTU's per SCF heating value. The Texas Gulf cluster model is the only one that did not use all the available natural gas. Since total refinery fuel usage for this category represented 12.1% of crude intake (exclusive of purchased electricity and catalytic cracking catalyst coke), this cluster is not typical when compared to the rest of the U. S. Since the Texas Gulf Coast refineries have extensive specialty (i.e., lube processing) and petrochemicals operations which are not simulated by the ADL refinery model, we felt that it was not required that a close calibration be achieved on total refinery fuel usage and purchased natural gas. On an FOE basis, the difference in refinery fuel consumption (13.6 MB/CD) is similar to the difference in purchased natural gas (15.8 MB/CD), indicating that internal fuel generation and consumption is in balance. Electricity consumption (purchased and internally generated) is also shown in the energy consumption category.

Finally, in Table I-18, is shown a comparison entitled "Processing Summary" presenting the various intakes/severities of the model run results with data reported by industry and capacity data obtained from the Oil and Gas Journal. We do not know if all the data reported by industry on catalytic reforming severity (RON) includes those reformers

Table I-17. LOUISIANA GULF CLUSTER MODEL
Calibration Results
(MB/CD)[a]

		Oil and Gas capacity MB/SD	BOM data	Industry data	Model run
Material balances					
Total crude intake			219.8	216.8	222.2
Energy consumption					
Purchased natural gas (FOE)			5.4	–	5.4
Total fuel consumption (FOE)[b]			17.2	–	17.0
Electricity MKWH/CD			606	–	744
Processing summary					
Catalytic reforming	intake	40.5		32.9	28.3
	severity RON	–		92.3	90.0
Catalytic cracking	intake	95.0		90.0	82.2
	conversion % vol.	–		70.0	69.6
Alkylation	production	24.0		17.7	17.5
Hydrocracking	intake	9.5		6.3	6.6
Coking	intake	20.0		19.1	15.8

[a] MB/CD unless otherwise noted.

[b] Excludes catalyst coke.

Table I-18. TEXAS GULF CLUSTER MODEL

Calibration Results

(MB/CD)[a]

		Oil and Gas capacity MB/SD	BOM data	Industry data	Model run
Material balances					
Total crude intake			325.9	331.4	331.4
Energy consumption					
Purchased natural gas (FOE)			29.2	–	13.4
Total fuel consumption (FOE)[b]			40.2	–	26.6
Electricity MKWH/CD			1078	–	1300
Processing summary					
Catalytic reforming	intake	82.3		73.8	70.6
	severity RON	–		94.8	90.0
Catalytic cracking	intake	114.8		100.4	92.7
	conversion % vol.	–		78.6	68.1
Alkylation	production	20.8		17.8	17.8
Hydrocracking	intake	21.3		14.7	14.7
Coking	intake	21.0		19.3	17.3
Isomerization	intake	2.4		2.4	0.0

[a] (MB/CD) unless otherwise noted.

[b] Excludes catalyst coke.

Table I-19. LARGE MIDWEST CLUSTER MODEL
Calibration Results
(MB/CD)[a]

		Oil and Gas capacity MB/SD	BOM data	Industry data	Model run
Material balances					
Total crude intake			146.1	145.6	145.6
Energy consumption					
Purchased natural gas (FOE)			.2	–	.2
Total fuel consumption (FOE)[b]			8.1	–	8.4
Electricity MKWH/CD			843	–	546
Processing summary					
Catalytic reforming	intake	32.7		27.8	27.6
	severity RON	–		90.7	90.0
Catalytic cracking	intake	55.0		51.2	48.7
	conversion % vol.	–		74.9	74.3
Alkylation	production	13.4		11.4	12.0
Coking	intake	15.8		13.6	14.1

[a] MB/CD unless otherwise noted.

[b] Excludes catalyst coke.

Table I-20. SMALL MIDCONTINENT CLUSTER MODEL

Calibration Results

(MB/CD)[a]

		Oil and Gas capacity MB/SD	BOM data	Industry data	Model run
Material balances					
Total crude intake			56.1	56.6	55.1
Energy consumption					
Purchased natural gas (FOE)			2.0	–	2.0
Total fuel consumption (FOE)[b]			4.4	–	4.7
Electricity MKWH/CD			284	–	213
Processing summary					
Catalytic reforming	intake	15.5		13.3	14.5
	severity RON	–		85.8	91.4
Catalytic cracking	intake	22.8		20.2	19.5
	conversion % vol.	–		76.5	77.3
Alkylation	production	5.3		5.0	4.9
Coking	intake	4.6		4.6	4.3
Isomerization	intake	1.8		1.4	–

[a] MB/CD unless otherwise noted.

[b] Excludes catalyst coke.

Table I-21. WEST COAST CLUSTER MODEL

Calibration Results

(MB/CD)[a]

		Oil and Gas capacity MB/SD	BOM data	Industry data	Model run
Material balances					
Total crude intake			159.6	162.0	155.2
Energy consumption					
Purchased natural gas (FOE)			3.4	–	6.4
Total fuel consumption (FOE)[b]			14.0	–	14.8
Electricity MKWH/CD			1262	–	768
Processing summary					
Catalytic reforming	intake	43.3		34.7	37.3
	severity RON	–		95.9	92.6
Catalytic cracking	intake	51.6		44.5	35.0
	conversion % vol.	–		61.6	61.0
Alkylation	production	7.8		5.6	5.5
Hydrocracking	intake	27.8		23.2	22.1
Coking	intake	42.1		38.4	39.4

[a] MB/CD unless otherwise noted.

[b] Excludes catalyst coke.

Table I-22. EAST COAST CLUSTER MODEL

Calibration Results

(MB/CD)[a]

		Oil and Gas capacity MB/SD	BOM data	Industry data	Model run
Material balances					
Total crude intake			187.4	186.5	188.0
Energy consumption					
Purchased natural gas (FOE)			2.3	–	2.5
Total fuel consumption (FOE)[b]			13.9	–	12.6
Electricity MKWH/CD			1365	–	712
Processing summary					
Catalytic reforming	intake	49.3		40.7	39.5
	severity RON	–		93.4	95.4
Catalytic cracking	intake	71.8		73.1	72.4
	conversion % vol.	–		71.6	67.5
Alkylation	production	8.4		6.9	8.0
Hydrocracking	intake	10.0		7.4	7.5

[a] MB/CD unless otherwise noted.

[b] Excludes catalyst coke

operating primarily for BTX production. The reported ADL reforming severity is for motor gasoline blending use only.

The West Coast and East Coast cluster model runs (Tables I-21 and I-22) consumed more natural gas than was indicated to be available by the BOM data. In these cases we increased the allowable maximum above BOM statistics to account for hydrogen plant feedstock use. The Louisiana and Texas model runs consumed more electricity than indicated by BOM data while the other four consumed less.

The Large Midwest cluster (Table I-19) contains two refineries built since 1970. Since these refineries are among the most modern in the U. S. it is expected that their fuel efficiency will be much greater than the U. S. average. It was decided to reduce the average unit fuel consumptions outlined in Appendix H to 80% of the usual levels to improve calibration. After this adjustment was made, the refinery material balance and energy consumption checked quite well against BOM data; this adjustment was maintained for all future runs.

Tables I-23 through I-28 present gasoline blending summaries for each cluster model which include a comparison of gasoline average sulfur content, clear pool octanes, and lead additions with industry data. For example, Table I-24 contains the gasoline blending summary for the Texas Gulf. Included in the table is a tabulation of each blending component and the respective volumes routed to premium, regular and low lead gasolines. The sulfur content (PPM) is shown for each blending component and the average for the pool was calculated to be 524 PPM compared with the reported industry data of 410 PPM. The model run required 1.99 grams per gallon of lead addition compared to the industry data of 2.09.

In general, all cluster model results checked well against industry data in regard to pool octanes and lead addition. Sulfur contents also checked quite well with the exception of the Louisiana Gulf Coast and East Coast. For the Louisiana Gulf Coast the industry reported an average of 440 PPM which appears to us to be unreasonably high considering the average sulfur content of the crude slate to this cluster model. Conversely, the industry reported East Coast value of 230 PPM appears to be low by a factor of 2-3.

Table I-23. LOUISIANA GULF CALIBRATION

Gasoline Blending Summary

Stream/Quality	Gasoline blend components, MB/CD				Component stream sulfur, PPM	Gasoline pool qualities	
	Premium	Regular	Low lead	Total pool		Model run	Industry data
Component stream							
90 RON reformate	–	21.40	3.78	25.18	1		
Cat cracker gasoline	21.34	19.46	3.48	44.28	676		
Normal butane	4.14	4.95	.77	9.86	1		
Alkylate	15.56	1.13	.76	17.45	3		
Straight run	–	13.65	–	13.65	72		
Light hydrocrackate	–	1.62	–	1.62	1		
Coker gasoline	–	1.66	–	1.66	2359		
Desulf. coker gasoline	–	1.83	–	1.83	4		
Natural gasoline	–	3.25	–	3.25	20		
Total pool	41.04	68.95	8.79	118.78			
Gasoline pool qualities							
Sulfur, PPM						295	440
RON clear						88.0	89.0
MON clear						81.2	81.6
Lead addition gm/gal						1.83	1.77

Table I-24. TEXAS GULF CALIBRATION

Gasoline Blending Summary

Stream/Quality	Gasoline blend components, MB/CD				Component stream sulfur, PPM	Gasoline pool qualities	
	Premium	Regular	Low lead	Total pool		Model run	Industry data
Component stream							
90 RON reformate	3.08	43.00	–	46.08	1		
Cat cracker gasoline	20.80	17.79	10.76	49.35	1462		
Normal butane	4.26	5.81	–	10.07	1		
Alkylate	14.01	–	3.81	17.82	3		
Straight run	.34	14.30	–	14.64	160		
Light hydrocrackate	1.69	–	3.43	5.12	1		
Coker gasoline	–	1.72	–	1.72	4161		
Desulf. coker gasoline	–	.10	–	.10	4		
Natural gasoline	2.95	8.11	–	11.06	16		
BTX raffinate	–	9.66	–	9.66	1		
Total pool	47.13	100.49	18.00	165.62			
Gasoline pool qualities							
Sulfur, PPM						524	410
RON clear						88.1	88.2
MON clear						79.9	79.7
Lead addition gm/gal						1.99	2.09

Table I-25. SMALL MIDCONTINENT CALIBRATION

Gasoline Blending Summary

Stream/Quality	Gasoline blend components, MB/CD				Component stream sulfur, PPM	Gasoline pool qualities	
	Premium	Regular	Low lead	Total pool		Model run	Industry data
Component stream							
90 RON reformate	–	8.08	–	8.08	1		
100 RON reformate	–	.74	.55	1.29	1		
Cat cracker gasoline	3.74	6.97	–	10.71	738		
Normal butane	.64	1.96	.04	2.64	1		
Isobutane	–	–	.01	.01	1		
Alkylate	2.51	2.34	.06	4.91	3		
Straight run	.86	2.93	–	3.79	103		
Coker gasoline	–	.42	–	.42	1886		
Natural gasoline	–	3.26	.70	3.96	20		
BTX raffinate	–	1.64	–	1.64	1		
Total pool	7.75	28.34	1.36	37.45			
Gasoline pool qualities							
Sulfur, PPM						243	340
RON clear						87.1	86.1
MON clear						81.2	80.0
Lead addition gm/gal						1.41	1.58

Table I-26. LARGE MIDWEST CALIBRATION

Gasoline Blending Summary

Stream/Quality	Gasoline blend components, MB/CD				Component stream sulfur, PPM	Gasoline pool qualities	
	Premium	Regular	Low lead	Total pool		Model run	Industry data
Component stream							
90 RON reformate	–	23.09	–	23.09	1		
Cat cracker gasoline	8.57	18.51	.08	27.16	2265		
Normal butane	1.66	5.17	.02	6.85	1		
Alkylate	6.27	5.67	.07	12.01	1		
Straight run	–	9.54	–	9.54	284		
Coker gasoline	–	1.40	–	1.40	4896		
Natural gasoline	–	.71	–	.71	20		
BTX raffinate	–	.94	–	.94	1		
Total pool	16.50	65.03	.17	81.70			
Gasoline pool qualities							
Sulfur, PPM						843	820
RON clear						88.5	87.9
MON clear						81.2	79.6
Lead addition gm/gal						1.38	1.74

Table I-27. WEST COAST CALIBRATION

Gasoline Blending Summary

Stream/Quality	Gasoline blend components, MB/CD				Component stream sulfur, PPM	Gasoline pool qualities	
	Premium	Regular	Low lead	Total pool		Model run	Industry data
Component stream							
100 RON reformate	–	–	.25	.25	1		
95 RON reformate	11.08	–	–	11.08	1		
90 RON reformate	–	11.07	–	11.07	1		
Cat cracker gasoline	14.49	4.06	–	18.55	2047		
Normal butane	3.10	.97	.01	4.08	1		
Alkylate	5.53	–	–	5.53	3		
Straight run	–	5.60	.07	5.67	203		
Light hydrocrackate	2.32	2.59	.20	5.11	1		
Coker gasoline	–	4.14	–	4.14	4613		
Natural gasoline	–	.81	–	.81	20		
BTX raffinate	–	1.49	–	1.49	1		
Total pool	36.52	30.73	.53	67.78			
Gasoline pool qualities							
Sulfur, PPM						860	700
RON clear						89.0	90.4
MON clear						80.3	81.5
Lead addition gm/gal						2.11	1.94

Table I-28. EAST COAST CALIBRATION

Gasoline Blending Summary

Stream/Quality	Gasoline blend components, MB/CD				Component stream sulfur, PPM	Gasoline pool qualities	
	Premium	Regular	Low lead	Total pool		Model run	Industry data
Component Stream							
100 RON reformate	13.87	–	1.00	14.87	1		
95 RON reformate	–	–	4.61	4.61	1		
90 RON reformate	–	11.24	–	11.24	1		
Cat cracker gasoline	5.39	34.23	–	39.62	1447		
Normal butane	2.97	4.33	.55	7.85	1		
Alkylate	7.98	–	–	7.98	3		
Straight run	–	10.72	–	10.72	79		
Light hydrocrackate	–	1.77	.45	2.22	1		
Natural gasoline	–	4.44	–	4.44	20		
BTX raffinate	–	1.36	–	1.36	1		
Total pool	30.21	68.09	6.61	104.91			
Gasoline pool qualities							
Sulfur, PPM						556	230
RON clear						89.5	87.8
MON clear						81.8	80.0
Lead addition gm/gal						1.60	1.82

In order to improve the octane calibration, the basic gasoline blending values shown in Table H-12 in Appendix H were modified somewhat for most cluster models. For the Louisiana Gulf, the Large Midwest, Small Midcontinent and East Coast all leaded research octanes were lowered 0.5 units and leaded motor octane numbers were increased 1.0 units. The leaded motor octane numbers only were increased by 0.2 units for the Texas Gulf Coast and no octanes were changed for the West Coast.

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APPENDIX J

STUDY RESULTS

This appendix gives a detailed summary of the results of this study. Tables J-1 through J-4 give the economic and energy penalties for the total U.S. refining industry for the reduction of refinery sulfur oxides emissions. These have been calculated by scaling up the LP model results using the scale up factors derived in Appendix G. The scaled up capital investment requirements and operating costs used in evaluating the economic penalties are given in Tables J-5 and J-6. Table J-7 provides existing capacity and calibration utilization of capacity for reforming, hydrocracking, alkylation and isomerization. These figures were used as the basis for determining capital requirements, as discussed in Appendix E.

The LP model results are given in Tables J-8 through J-47. Capital investments and operating costs for the LP model runs are given in Tables J-8 through J-15. Crude slates and other inputs, product outputs, and process unit throughput and severities are given in Tables J-16 through J-29. Gasoline blends are given in Tables J-30 through J-43. The volumes of refinery fuel consumed and the amount of residual fuel oil produced, together with their sulfur levels are given in Tables J-44 through J-47.

A. MASS AND SULFUR BALANCE

The computer model operates on a volumetric basis, and each process unit yield structure was carefully checked to ensure that the volumetric process outturns were reliable. The sulfur content of each stream was entered in lbs./barrel, and was similarly checked and mass balanced for specific feedstocks. At the completion of the study, it was deemed to be desirable to illustrate the method of checking overall mass balances and sulfur balances, for the benefit of interested parties. Hence, sample calculations are provided herein to illustrate this procedure.

To complete these sample calculations, specific gravities were assumed for every stream in the refinery simulation. Using the stream flow rates from the computer output and the assumed gravities, the weight balances shown in Tables J-54 and J-55 were constructed. Because of the time which would be required to refine these assumed gravities, the input and output streams do not balance precisely; hence, the results illustrate the method of calculation, but are not indicative of the actual mass balances in the simulation.

Figure J-1 and Tables J-54 and J-55 detail the hydrocarbon and sulfur flows by individual process units for the Texas Gulf cluster, with numbered arrows on the flowsheet corresponding to stream numbers and stream names listed on the tables. Table J-54 gives volume, mass, and sulfur flows for Scenario B/C and Table J-55 for Scenario F. Refinery streams of C_4 and lighter, as well as gases, such as H_2S and SO_x, are grouped in verticle output arrows on the flowsheet and in streams labeled "light ends" on the tables.

Following is a discussion of the methodology used in the Texas Gulf mass and sulfur balance for 1985, Scenario B/C.

1. Crude-Specific Streams

Stream values for crude-specific streams are calculated from information on process intakes (Tables J-23 through J-29), yield on crude or process yields, hydrogen consumption, sulfur removal in desulfurization and stream qualities (Appendix H).

For example, arrow #5 on Figure J-1 represents the aggregated flow of gas oils 375-650°F for the crudes charged to the atmospheric distillation tower. The unaggregated values for each crude's gas oil is calculated and summed in Table J-48. The crude volume or charge (column 1) multiplied by the yield on crude (column 2) gives the stream volume (column 5). The stream volume multiplied by the stream density gives the hydrocarbon weight for that stream. Sulfur content (column 4) divided by 100 and multiplied by the hydrocarbon weight yields the sulfur weight (column 7). Sulfur in PPM (parts per million) is derived from the hydrocarbon and sulfur weights as shown. Stream values for other crude-specific streams are calculated in a similar manner with the exception of reformate whose qualities for light, medium and heavy straight-run naphtha are given in Appendix H.

Desulfurization of crude-specific streams requires additional information on hydrogen consumption and the level of sulfur removal. For example, desulfurization of light gas oil requires 190 SCF of hydrogen per barrel of intake and converts 99% of the feed sulfur into H_2S, leaving the remaining 1% in the liquid output stream (see Table J-49).

Isomerization takes in desulfurized C_5 to 160°F for all crudes except Nigerian, whose undesulfurized stream is at the required 1 PPM sulfur level.

2. Cluster-Specific Streams

Cluster-specific streams, output streams of the catalytic cracker, coker and visbreaker, have constant specific gravities (with the exception of desulfurized FCC feed) and sulfur contents which vary according to the feed sulfur level (i.e., with the crude slate and cluster). The feed sulfur to each process unit is distributed among the products according to the percentages given in Appendix H. Because the actual feed to these units cannot be known until after the LP solution has been reached, an estimate of each unit's feed is made prior to LP optimization in order to set output stream sulfur contents. Catalytic cracker feed is assumed to be vacuum overhead (650-1050°F) and coker/visbreaker feed is assumed to be vacuum bottoms (1050°F+). Each crude is assumed to be represented by a factor equal to its percentage of the crude slate times its yield of the specified feed stream. The hydrocracker has one cluster-specific stream, H_2S, which accumulates all feed sulfur not contained in the unit's liquid output streams. Feed to the hydrocracker is assumed to be heavy gas oil (500-650°F). Table J-50 shows assumed feed sulfur levels for the catalytic cracker, visbreaker/coker and hydrocracker. Table J-51 distributes this feed sulfur among the process output streams and lists the stream qualities and output of H_2S and SO_x.

The cluster models allow greater flexibility in feed streams to the conversion units than the assumed feeds discussed above. Each unit takes in hydrocarbons within the specified boiling range yet is not limited to either a fixed ratio of crudes in straight-run streams nor to straight-run streams alone. This feed flexibility and the necessity of assuming feed sulfur in order to set product sulfur levels is a potential source of error in the sulfur balance around each conversion unit.

3. Miscellaneous Streams

Miscellaneous streams are handled in a manner similiar to the crude-specific streams, using yield data and stream qualities.

Densities used for refinery gas, BTX, olefins, coke and hydrogen are shown in Table J-53.

Table J-1. ECONOMIC PENALTY FOR REDUCING REFINERY SO_x EMISSIONS[a] – 1977

Scenario F Versus Scenario C

	East Coast	Large Midwest	Small Midcontinent	Louisiana Gulf	Texas Gulf	East of Rockies Grassroots	Subtotal PAD I-IV	West Coast	West of Rockies Grassroots	Subtotal PAD V	Total U.S.A.
Basis											
Cumulative capital investment required millions $	397.1	1016.7	414.9	304.3	598.8	–	2731.8	388.8	–	388.8	3120.6
Additional crude oil processed MB/CD	1.0	5.0	22.3	15.9	–	–	44.2	0.4	–	0.4	44.6
Additional LPG produced MB/CD	(3.4)	1.4	2.6	2.5	(4.3)	–	–	(0.5)	–	(0.5)	(1.7)
Penalties											
Thousands dollars per day											
Capital charge (25%)[a]	272	696	284	208	410	–	1870	266	–	266	2136
Operating costs	65	140	30	31	73	–	339	54	–	54	393
Crude oil @ $12.5/B	13	63	279	199	–	–	554	5	–	5	559
LPG @ $8.75/B	30	(12)	(23)	(22)	38	–	11	4	–	4	15
Sulfur @ $10/ton	(6)	(3)	(1)	(1)	(5)	–	(16)	(3)	–	(3)	(19)
Total penalty							2758			326	3084
Total products MB/CD	1685	2630	1026	1820	3873	–	11034	2361	–	2361	13395
Penalty $/B of total products[a]							0.25			0.14	0.23
Penalty ¢/G of total products[a]							0.60			0.33	0.56

---

[a] Based on cumulative capital investment.

Table J-2. ECONOMIC PENALTY FOR REDUCING REFINERY SO_x EMISSIONS[a] – 1985

Scenario F Versus Scenario C

	East Coast	Large Midwest	Small Midcontinent	Louisiana Gulf	Texas Gulf	East of Rockies Grassroots	Subtotal PAD I-IV	West Coast	West of Rockies Grassroots	Subtotal PAD V	Total U.S.A.
Basis											
Cumulative capital investment required millions $	435.4	967.9	347.7	311.7	861.7	873.9	3798.3	420.2	274.9	695.1	4493.4
Additional crude oil processed MB/CD	–	–	–	–	–	47.4	47.4	–	15.2	15.2	62.6
Additional LPG produced MB/CD	(4.0)	3.2	(1.6)	(32.6)	(3.8)	–	(38.8)	(0.5)	–	(0.5)	(39.3)
Penalties											
Thousands dollars per day											
Capital charge (25%)[a]	298	663	238	214	590	599	2602	288	188	476	3078
Operating costs	62	114	25	38	96	123	458	50	66	116	574
Crude oil @ $12.5/B	–	–	–	–	–	593	593	–	190	190	783
LPG @ $8.75/B	35	(27)	14	285	33	–	340	4	–	4	344
Sulfur @ $10/ton	(5)	(4)	(1)	(1)	(6)	(6)	(23)	(2)	(1)	(3)	(26)
Total penalty							3970			783	4753
Total products MB/CD	1631	2566	924	1703	3633	2787	13234	1996	598	2594	15828
Penalty $/B of total products[a]							0.30			0.30	0.30
Penalty ¢/G of total products[a]							0.71			0.72	0.71

---

[a] Based on cumulative capital investment.

Table J-3. ENERGY PENALTY FOR REDUCING REFINERY SO_x EMISSIONS – 1977

Scenario F Versus Scenario C

	East Coast	Large Midwest	Small Midcontinent	Louisiana Gulf	Texas Gulf	West Coast	Total U.S.A.
Basis							
Additional crude oil processed MB/CD	1.0	5.0	22.3	15.9	–	0.4	44.6
Additional LPG produced MB/CD	(3.4)	1.4	2.6	2.5	(4.3)	(0.5)	(1.7)
Additional purchased power required MKWH/CD	1292	1594	443	524	1294	1199	6346
Energy penalties 10^9 BTU/CD							
Crude oil	6	28	125	89	–	2	250
LPG	14	(5)	(10)	(10)	17	2	8
Purchased power	13	16	4	5	13	12	63
Total penalty 10^9 BTU/CD							321
Thousands barrels of fuel oil equivalent per calendar day							51.0

Table J-4. ENERGY PENALTY FOR REDUCING REFINERY SO_x EMISSIONS – 1985

Scenario F Versus Scenario C

	East Coast	Large Midwest	Small Midcontinent	Louisiana Gulf	Texas Gulf	East of Rockies Grassroots	Subtotal PAD I-IV	West Coast	West of Rockies Grassroots	Subtotal PAD V	Total U.S.A.
Basis											
Additional crude oil processed MB/CD	–	–	–	–	–	47.4	47.4	–	15.2	15.2	62.6
Additional LPG produced MB/CD	(4.0)	3.2	(1.6)	(32.6)	(3.8)	–	(38.8)	(0.5)	–	(0.5)	(39.3)
Additional purchased power required MKWH/CD	1134	1831	442	484	1783	2217	7891	900	1324	2224	10115
Energy penalties 10^9 BTU/CD											
Crude oil	–	–	–	–	–	265	265	–	85	85	350
LPG	16	(13)	6	131	15	–	155	2	–	2	157
Purchased power	11	18	4	5	18	23	79	9	13	22	101
Total penalty 10^9 BTU/CD							499			109	608
Thousands barrels of fuel oil equivalent per calendar day							79.2			17.3	96.5

Table J-5. CAPITAL INVESTMENT REQUIREMENTS TO REDUCE REFINERY SOX EMISSION LEVELS

Million $ 1st Quarter 1975

Cluster	Scenario F versus C						Sox emission levels 1985 Short tons per day				
	Cluster investment			Scaled up investment			Cluster levels		Scaled up levels		
	1977	1985	Total	1977	1985	Total	before control	after control	before control	after control	percent reduction
East Coast	43.7	4.3	48.0	397.1	38.3	435.4	59	14	449	107	76
Large Midwest	47.1	(2.3)	44.8	1,016.7	(48.8)	967.9	73	14	1,323	254	81
Small Midcontinent	21.2	(3.5)	17.7	414.9	(67.2)	347.7	24	6	393	98	75
Louisiana Gulf	32.6	0.8	33.4	304.3	7.4	311.7	24	3	188	23	88
Texas Gulf	45.3	20.3	65.6	598.8	262.9	861.7	92	35	1,019	387	62
West Coast	24.2	2.2	26.4	388.8	31.4	420.2	47	13	572	158	72
West Grassroots				–	274.9	274.9	55	19	165	57	65
East Grassroots				–	873.9	873.9	89	16	1,335	240	82
Total				3,120.6	1,372.8	4,493.4			5,444	1,324	76

Table J-6. OPERATING COSTS REQUIRED TO REDUCE REFINERY SOX EMISSION LEVELS

(thousands $ per day)

Cluster	Scenario F versus C			
	Cluster operating cost		Scaled up operating cost	
	1977	1985	1977	1985
East Coast	8.4	8.1	64.8	61.6
Large Midwest	7.6	6.3	139.6	114.2
Small Midcontinent	1.8	1.5	29.9	24.6
Louisiana Gulf	3.9	4.8	30.9	37.5
Texas Gulf	6.5	8.7	72.9	96.3
West Coast	3.9	4.1	53.5	49.9
West Grassroots		22.1		66.3
East Grassroots		8.2		123.0
Total			391.6	573.4

Table J-7. BASIS FOR CLUSTER CAPITAL INVESTMENT REQUIREMENTS

Existing Capacity Versus Calibration Run Requirement

	East Coast	Large Midwest	Small Midcont.	Louisiana Gulf	Texas Gulf	West Coast
Process (MB/CD)						
Reforming						
Existing capacity for BTX	3.5	2.3	4.3	–	20.3	16.3
Existing capacity for mogas – high severity	7.0	4.6	–	8.6	–	–
– low severity	31.[illegible]	20.9	8.9	25.8	49.7	20.5
1973 Calibration utilization – high severity	7.0	–	–	–	–	–
– low severity	29.0	25.3	10.2	28.3	50.4	21.0
Spare capacity available	2.4	0.2	–	6.1	–	–
Hydrocracking						
Existing capacity – high severity	7.2	–	–	–	18.1	–
– medium severity	1.3	–	–	8.1	–	23.6
1973 Calibration utilization – high	6.2	–	–	–	14.7	–
– medium	1.3	–	–	6.6	–	22.1
Spare capacity available	1.0	–	–	1.5	3.4	1.5
Alkylation						
Existing capacity	7.1	11.4	4.5	20.4	17.7	6.6
1973 Calibration utilization	8.0	12.0	4.9	17.5	17.8	5.5
Spare capacity available	–	–	–	2.9	–	1.1
Isomerization						
Existing capacity – once through	–	–	1.5	–	2.0	–
1973 Calibration utilization	–	–	–	–	–	–
Spare capacity available	–	–	1.5	–	2.0	–

Table J-8. L.P. MODEL RESULTS: – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: East Coast		Scenario: C				Scenario: F		
		1977	1980	1985	Total	1977	1985	Total
Capital investments	(million dollars) (1st Q 1975 Basis)							
Reforming:	existing capacity	1.6	–	–	1.6	1.6	–	1.6
	severity upgrade	11.9	1.8	6.0	19.7	12.3	6.9	19.8
	new capacity	4.5	–	–	4.5	4.5	–	4.5
Hydrocracking:	existing capacity	1.1	–	–	1.1	1.1	–	1.1
	new capacity	0.7	0.1	0.1	0.9	1.7	–	1.7
Isomerization:	once through upgrading	–	–	1.9	1.9	–	–	–
	new capacity	0.1	2.1	5.3	7.5	1.2	10.4	11.6
Alkylation:	new capacity	4.8	1.7	0.4	6.9	1.4	2.1	3.5
Light naphtha desulfurization: new capacity		–	1.1	1.4	2.5	0.3	3.0	3.3
Cat cracker feed desulfurization: new capacity		–	–	–	–	21.9	2.1	24.0
Sulfur recovery		–	–	–	–	4.6	–	4.6
Subtotals		24.7	6.8	15.1	46.6	51.2	24.5	75.7
Offsites and working capital at 40%		9.9	2.7	6.0	18.6	20.5	9.8	30.3
Subtotals		34.6	9.5	21.1	65.2	71.7	34.3	106.0
Totals (adjusted for stream day)		40.7	11.2	24.8	76.7	84.4	40.3	124.7
Operating costs (thousands of dollars per day)								
Purchased steam		5.7	5.7	5.7		5.7	5.7	
Electricity		16.8	17.1	18.0		20.2	21.2	
Cooling water		15.9	16.2	16.6		16.8	17.3	
Maintenance		28.3	29.1	30.8		31.4	34.2	
Manpower		53.1	53.3	53.8		54.0	54.8	
Tetra ethyl lead		8.6	4.3	–		8.6	–	
Catalysts and chemicals		17.3	18.4	20.1		17.4	19.9	
Totals		145.7	144.1	145.0		154.1	153.1	

Table J-9. L.P. MODEL RESULTS: – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: Large Midwest	Scenario: C				Scenario: F		
	1977	1980	1985	Total	1977	1985	Total
Capital investments (million dollars) (1st Q 1975 Basis)							
Reforming: existing capacity	0.1	–	–	0.1	0.1	–	0.1
severity upgrade	7.4	5.9	0.4	13.7	13.7	–	13.7
new capacity	1.4	–	1.4	2.8	0.7	5.1	5.8
Isomerization: new capacity	–	–	8.7	8.7	–	9.3	9.3
Alkylation: new capacity	–	1.1	–	1.1	–	–	–
Light naphtha desulfurization: new capacity	–	–	2.2	2.2	–	2.4	2.4
Cat cracker feed desulfurization: new capacity	–	–	–	–	15.3	0.2	15.5
Hydrogen manufacture: new capacity	–	–	–	–	3.3	0.4	3.7
Sulfur recovery	–	–	–	–	4.4	0.8	5.2
Subtotals	8.9	7.0	12.7	28.6	37.5	18.2	56.7
Offsites and working capital at 40%	3.6	2.8	5.0	11.4	15.0	7.3	22.3
Subtotals	12.5	9.8	17.7	40.0	52.5	26.5	78.0
Totals (adjusted for stream day)	14.7	11.5	20.8	47.0	61.8	30.0	91.8
Operating costs (thousands of dollars per day)							
Electricity	9.8	10.3	11.1		11.4	12.9	
Cooling water	11.3	11.9	12.0		11.7	12.1	
Maintenance	21.1	21.9	23.3		24.5	26.4	
Manpower	38.3	38.5	38.9		39.3	39.8	
Tetra ethyl lead	7.8	3.8	–		7.6	–	
Catalysts and chemicals	10.3	12.5	13.3		11.7	13.7	
Totals	98.6	98.9	98.6		108.2	104.9	

Table J-10. L.P. MODEL RESULTS: – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: Small Midcontinent		Scenario: C				Scenario: F		
		1977	1980	1985	Total	1977	1985	Total
Capital investments (million dollars) (1st Q 1975 Basis)								
Reforming:	severity upgrade	–	7.9	1.6	9.5	1.8	8.3	10.1
	new capacity	2.2	–	–	2.2	2.0	0.1	2.1
Isomerization:	existing capacity	–	–	1.4	1.4	0.9	0.6	1.5
	once through upgrading	–	–	–	–	0.9	–	0.9
	new capacity	1.0	2.2	3.2	6.4	–	5.3	5.3
Alkylation:	new capacity	0.2	0.2	0.7	1.1	–	–	–
Light naphtha desulfurization: new capacity		–	–	1.6	1.6	–	1.5	1.5
Cat cracker feed desulfurization: new capacity		–	–	–	–	9.6	0.6	10.2
Sulfur recovery		–	–	–	–	1.1	0.3	1.4
Subtotals		3.4	10.3	8.5	22.2	16.3	16.7	33.0
Offsites and working capital at 40%		1.4	4.1	3.4	8.9	6.5	6.7	13.2
Subtotals		4.8	14.4	11.9	31.1	22.8	23.4	46.2
Totals (adjusted for stream day)		5.6	17.0	14.0	36.6	26.8	27.5	54.3
Operating costs (thousands of dollars per day)								
Electricity		3.9	4.1	4.4		4.3	4.9	
Cooling water		4.5	4.5	4.7		4.4	4.7	
Maintenance		7.9	9.1	9.6		9.4	10.8	
Manpower		20.9	21.2	21.4		21.3	21.7	
Tetra ethyl lead		2.9	1.4	–		2.8	–	
Catalysts and chemicals		4.8	5.5	6.1		4.5	5.6	
Totals		44.9	45.8	46.2		46.7	47.7	

Table J-11. L.P. MODEL RESULTS: – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: Louisiana Gulf		Scenario: C				Scenario: F		
		1977	1980	1985	Total	1977	1985	Total
Capital investments	(million dollars) (1st Q 1975 Basis)							
Reforming:	existing capacity	1.3	0.9	1.8	4.0	4.0	–	4.0
	severity upgrade	4.5	10.6	1.8	16.9	2.6	14.3	16.9
	new capacity	–	–	0.8	0.8	2.0	1.4	3.4
Hydrocracking:	existing capacity	1.6	–	–	1.6	1.6	–	1.6
	severity change	–	–	–	–	1.2	–	1.2
	new capacity	0.8	–	–	0.8	0.5	0.2	0.7
Isomerization:	new capacity	–	–	7.3	7.3	–	5.8	5.8
Alkylation:	existing capacity	–	0.1	0.7	0.8	–	–	–
Light naphtha desulfurization: new capacity		–	–	2.6	2.6	–	2.5	2.5
Cat cracker feed desulfurization: new capacity		–	–	–	–	14.7	2.8	17.5
Sulfur recovery		–	–	–	–	1.8	0.1	1.9
Subtotals		8.2	11.6	15.0	34.8	28.4	27.1	55.5
Offsites and working capital at 40%		3.3	4.6	6.0	13.9	11.4	10.8	22.2
Subtotals		11.5	16.2	21.0	48.7	39.8	37.9	77.7
Totals (adjusted for stream day)		13.5	19.1	24.7	57.3	46.8	44.6	91.4
Operating costs (thousands of dollars per day)								
Electricity		11.6	12.0	12.3		12.5	13.2	
Cooling water		16.5	16.9	17.3		16.5	17.8	
Maintenance		28.4	29.7	31.5		30.7	33.9	
Manpower		43.6	44.0	44.5		44.3	45.2	
Tetra ethyl lead		11.7	5.7	–		11.5	–	
Catalysts and chemicals		13.1	14.5	16.6		13.4	16.9	
Totals		124.9	122.8	122.2		128.9	127.0	

Table J-12. L.P. MODEL RESULTS: – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: Texas Gulf		Scenario: C				Scenario: F		
		1977	1980	1985	Total	1977	1985	Total
Capital investments	(million dollars) (1st Q 1975 Basis)							
Reforming:	severity upgrade	30.7	1.3	1.0	33.0	23.6	9.4	33.0
	new capacity	0.8	1.7	0.9	3.4	–	2.7	2.7
Hydrocracking:	existing capacity	4.3	–	–	4.3	1.1	3.2	4.3
	severity change	3.0	–	0.6	3.6	3.7	0.6	4.3
	new capacity	2.0	–	0.8	2.8	–	2.3	2.3
Isomerization:	existing capacity	1.2	–	–	1.2	1.2	–	1.2
	once through upgrading	1.1	–	–	1.1	1.1	–	1.1
	new capacity	0.7	4.7	6.9	12.3	2.6	12.8	15.4
Alkylation:	new capacity	1.3	–	0.6	1.9	0.4	–	0.4
Light naphtha desulfurization: new capacity		0.2	1.2	2.8	4.2	0.7	3.8	4.5
Cat cracker feed desulfurization: new capacity		–	–	–	–	32.6	–	32.6
Sulfur recovery		–	–	–	–	5.8	–	5.8
Subtotals		45.3	8.9	13.6	67.8	72.8	34.8	107.6
Offsites and working capital at 40%		18.1	3.6	5.4	27.1	29.1	13.9	43.0
Subtotals		63.4	12.5	19.0	94.9	101.9	48.7	150.6
Totals (adjusted for stream day)		74.6	14.7	22.3	111.6	119.9	57.3	177.2
Operating costs (thousands of dollars per day)								
Electricity		19.7	20.0	20.2		21.3	22.5	
Cooling water		24.4	24.4	24.2		24.9	24.8	
Maintenance		47.5	48.5	48.3		50.7	52.9	
Manpower		114.9	115.2	115.1		115.8	116.5	
Tetra ethyl lead		15.9	7.8	–		15.9	–	
Catalysts and chemicals		26.1	26.6	28.7		26.4	28.5	
Totals		248.5	242.5	236.5		255.0	245.2	

Table J-13. L.P. MODEL RESULTS: – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: West Coast		Scenario: C				Scenario: F		
		1977	1980	1985	Total	1977	1985	Total
Capital investments (million dollars) (1st Q 1975 Basis)								
Reforming:	severity upgrade	5.2	3.4	5.2	13.8	5.0	7.5	12.5
	new capacity	0.8	–	0.1	0.9	1.5	–	1.5
Hydrocracking:	existing capacity	1.6	–	–	1.6	1.6	–	1.6
	new capacity	4.1	–	–	4.1	4.1	–	4.1
Isomerization:	new capacity	–	–	4.8	4.8	–	4.3	4.3
Alkylation:	existing capacity	1.5	–	–	1.5	1.5	–	1.5
	new capacity	1.8	–	0.4	2.2	1.1	1.1	2.2
Light naphtha desulfurization: new capacity		–	–	1.6	1.6	–	1.4	1.4
Cat cracker feed desulfurization: new capacity		–	–	–	–	9.4	2.6	12.0
Sulfur recovery		–	–	–	–	5.5	–	5.5
Subtotals		15.0	3.4	12.1	30.5	29.7	16.9	46.6
Offsites and working capital at 40%		6.0	1.4	4.8	12.2	11.9	6.7	18.6
Subtotals		21.0	4.8	16.9	42.7	41.6	23.6	65.2
Totals (adjusted for stream day)		24.7	5.6	19.9	50.2	48.9	27.8	76.7
Operating costs (thousands of dollars per day)								
Electricity		15.5	15.7	16.2		17.1	17.5	
Cooling water		12.1	12.3	12.5		12.3	12.8	
Maintenance		28.2	28.4	29.6		29.6	31.2	
Manpower		70.6	70.6	71.0		71.0	71.5	
Tetra ethyl lead		6.9	3.5	–		6.9	–	
Catalysts and chemicals		15.1	14.9	16.5		15.4	16.9	
Totals		148.4	145.4	145.8		152.3	149.9	

Table J-14. L.P. MODEL RESULTS: – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: Grassroots Refinery East of Rockies	Scenario C		Scenario F	
	sweet	sour	sweet	sour
Total capital investment (million dollars) (1st Q 1975 basis)	651.3	798.4	678.5	885.2
Operating costs (thousands dollars per day)				
Electricity	16.4	23.0	17.1	26.7
Cooling water	4.1	4.9	4.3	5.2
Maintenance	28.2	34.0	30.1	38.7
Manpower	22.7	34.1	23.3	35.5
Catalysts and chemicals	13.4	22.4	13.8	22.7
Total	84.8	118.4	88.6	128.8

Table J-15. L.P. MODEL RESULTS – CAPITAL INVESTMENT REQUIREMENTS AND OPERATING COSTS

Cluster: Grassroots Refinery – West of Rockies	Scenario C	Scenario F
Total capital investment (million dollars) (1st Q 1975 basis)	764.5	848.8
Operating costs (thousands dollars per day)		
Electricity	23.1	31.3
Cooling water	5.1	5.6
Maintenance	33.0	39.0
Manpower	33.4	35.1
Catalysts and chemicals	19.6	25.3
Total	114.2	136.3

Table J-16. L.P. MODEL RESULTS – FIXED INPUTS AND OUTPUTS

(MB/CD)

Cluster: East Coast	Year		
	1977	1980	1985
Inputs			
Tia Juana Medium crude oil	52.448	42.550	32.656
Saudi Arabian Light crude oil	70.656	80.550	90.447
Nigerian Forcados crude oil	34.042	36.020	38.000
Algerian Hassi Messoud crude oil	40.771	38.790	36.812
Isobutanes	0.315	0.280	0.245
Normal butanes	1.584	1.400	1.232
Natural gas (purchased refinery fuel)	1.875	1.250	–
Natural gasoline	5.840	5.840	5.840
Intermediate product transfer-cat feed	11.100	11.100	11.100
Intermediate product transfer-reformer feed	5.980	5.980	5.980
Total	224.611	223.760	222.312
Outputs			
Naphtha	1.333	1.328	1.320
Jet Fuel	6.001	5.978	5.941
Kerosine	3.532	3.518	3.497
Number 2 heating oil	47.431	47.249	46.954
Residual fuel oil	13.368	13.317	13.234
Lubes	5.147	5.127	5.096
Asphalt	18.891	18.818	18.700
BTX	1.417	1.411	1.403
Refinery gas	0.896	0.892	0.887
Olefin sales to petrochemicals	4.303	4.286	4.260
Total	102.319	101.924	101.292

Table J-17. L.P. MODEL RESULTS – FIXED INPUTS AND OUTPUTS

(MB/CD)

Cluster: Large Midwest	Year		
	1977	1980	1985
Inputs			
West Texas Sour crude oil	93.386	86.213	79.041
South Louisiana Mix crude oil	4.303	–	–
Oklahoma crude oil	6.886	6.599	6.312
Canadian Interprovincial mix crude oil	7.459	–	–
Saudi Arabian Light crude oil	31.416	50.638	58.097
Isobutanes	3.330	2.960	2.590
Natural gas (purchased refinery fuel)	0.180	0.120	–
Natural gasoline	0.837	0.744	0.651
Intermediate product transfer-cat feed	1.215	1.215	1.215
Intermediate product transfer-reformer feed	0.654	0.654	0.654
Total	149.666	149.143	148.560
Outputs			
Naphtha	2.122	2.115	2.107
Jet fuel	2.083	2.076	2.068
Kerosine	1.631	1.625	1.619
Number 2 heating oil	40.090	39.948	39.790
Residual fuel oil	7.261	7.236	7.207
Asphalt	3.744	3.730	3.716
Coke	3.616	3.603	3.589
BTX	0.924	0.920	0.917
Total	61.471	61.253	61.013

Table J-18. L.P. MODEL RESULTS – FIXED INPUTS AND OUTPUTS

(MB/CD)

Cluster: Small Midcontinent	Year		
	1977	1980	1985
Inputs			
West Texas Sour crude oil	6.593	6.043	5.494
South Louisiana Mix crude oil	3.296	1.648	–
Oklahoma crude oil	32.855	31.756	30.657
Canadian Interprovincial Mix crude oil	4.835	–	–
Saudi Arabian Light crude oil	3.681	7.747	9.395
Algerian Hassi Messoud crude oil	3.681	7.747	9.395
Isobutanes	0.846	0.752	0.658
Normal butanes	0.279	0.248	0.217
Natural gas (purchased refinery fuel)	1.535	1.023	–
Natural gasoline	4.950	4.400	3.850
Intermediate product transfer-cat feed	0.436	0.436	0.436
Intermediate product transfer-reformer feed	0.235	0.235	0.235
Total	63.222	62.035	60.337
Outputs			
Naphtha	0.343	0.336	0.327
Jet fuel	0.900	0.883	0.859
Kerosene	0.039	0.038	0.037
Number 2 heating oil	15.698	15.400	14.973
Residual fuel oil	0.225	0.221	0.215
Lubes	0.333	0.326	0.317
Asphalt	1.771	1.738	1.690
Coke	1.282	1.258	1.223
BTX	1.370	1.344	1.307
Refinery gas	0.528	0.518	0.504
Olefin sales to petrochemicals	0.587	0.576	0.560
Total	23.076	22.638	22.012

Table J-19. L.P. MODEL RESULTS – FIXED INPUTS AND OUTPUTS
(MB/CD)

Cluster: Louisiana Gulf	Year		
	1977	1980	1985
Inputs			
West Texas Sour crude oil	25.723	25.723	25.723
South Louisiana Mix crude oil	192.270	192.270	192.270
Isobutanes	5.490	4.880	4.700
Normal butanes	5.373	4.776	4.179
Natural gas (purchased refinery fuel)	4.050	2.700	–
Natural gasoline	3.852	3.424	3.000
Total	236.758	233.773	229.872
Outputs			
Naphtha	0.757	0.747	0.734
Jet fuel	17.984	17.757	17.427
Kerosine	5.338	5.271	5.173
Number 2 heating oil	65.801	64.972	63.783
Residual fuel oil	5.192	5.127	5.032
Asphalt	1.504	1.485	1.458
Coke	3.999	3.948	3.870
Refinery gas	0.311	0.307	0.301
Olefin sales to petrochemicals	2.329	2.300	2.257
Intermediate product transfer-cat feed	2.078	2.078	2.078
Intermediate product transfer-reformer feed	1.120	1.120	1.120
Total	106.413	105.112	103.213

Table J-20. L.P. MODEL RESULTS – FIXED INPUTS AND OUTPUTS

(MB/CD)

Cluster: Texas Gulf	Year		
	1977	1980	1985
Inputs			
West Texas Sour crude oil	135.964	135.964	135.964
South Louisiana Mix crude oil	155.669	155.669	155.669
Tia Juana Medium crude oil	6.897	6.897	6.897
Saudi Arabian Light crude oil	17.405	17.405	17.405
Nigerian Forcados crude oil	12.480	12.480	12.480
Isobutanes	1.917	1.704	1.491
Normal butanes	1.845	1.640	1.435
Natural gas (purchased refinery fuel)	10.085	6.724	–
Natural gasoline	14.400	12.800	11.200
Total	356.662	351.283	342.541
Outputs			
Naphtha	8.598	8.468	8.258
Jet fuel	22.952	22.604	22.043
Kerosene	7.524	7.410	7.228
Number 2 heating oil	81.383	80.131	78.141
Residual fuel oil	15.321	15.089	14.714
Lubes	16.112	15.868	15.474
Asphalt	1.368	1.347	1.314
Coke	3.906	3.849	3.754
BTX	5.911	5.822	5.677
Refinery gas	0.821	0.808	0.788
Olefin sales to petrochemicals	3.879	3.820	3.725
Intermediate product transfer-cat feed	1.954	1.925	1.877
Intermediate product transfer-reformer feed	4.842	4.768	4.650
Total	174.553	171.909	167.641

Table J-21. L.P. MODEL RESULTS – FIXED INPUTS AND OUTPUTS

(MB/CD)

Cluster: West Coast	Year		
	1977	1980	1985
Inputs			
California Ventura crude oil	21.673	21.673	21.673
California Wilmington crude oil	65.676	65.676	65.676
Alaskan North Slope crude oil	–	76.841	76.841
Canadian Interprovincial Mix crude oil	5.580	–	–
Saudi Arabian Light crude oil	54.840	–	–
Indonesian Minas crude oil	16.420	–	–
Isobutanes	0.450	0.400	0.350
Normal butanes	0.144	0.128	0.112
Natural gas (purchased refinery fuel)	4.793	3.195	–
Natural gasoline	1.170	1.040	0.910
Intermediate product transfer-cat feed	3.597	3.597	3.597
Intermediate product transfer-reformer feed	1.937	1.937	1.937
Total	176.280	174.487	171.096
Outputs			
Naphtha	3.978	3.936	3.900
Jet fuel	20.765	20.548	20.364
Kerosine	0.177	0.176	0.174
Number 2 heating oil	23.125	22.883	22.678
Residual fuel oil	31.894	31.561	31.279
Lubes	0.365	0.362	0.358
Asphalt	2.109	2.087	2.068
Coke – low sulfur	10.002	9.897	7.790
BTX	4.072	4.029	3.993
Refinery gas	0.480	0.475	0.470
Olefin sales to petrochemicals	1.357	1.343	1.331
Total	98.324	97.297	94.406

Table J-22. L.P. MODEL RESULTS: – INPUTS AND FIXED OUTPUTS

Grassroots Refineries

(MB/CD)

Inputs	East of Rockies			West of Rockies
	Sour crude refinery	Sweet crude refinery		
	Arabian Light crude oil	Nigerian crude oil	Algerian crude oil	Alaska North Slope crude oil
Scenario				
C	196.978	96.768	96.768	206.277
F	199.822	98.291	98.291	210.944

Fixed outputs	East of Rockies	West of Rockies
Jet fuel	9.700	31.200
Kerosine	2.900	–
Number 2 heating oil	46.900	30.900
Residual fuel oil	32.900	45.500
Total	92.400	107.600

Table J-23. L.P. MODEL RESULTS – PROCESSING AND VARIABLE OUTPUTS

East Coast Cluster

	Scenario C			Scenario F	
	1977	1980	1985	1977	1985
Variable output					
Gasoline MB/CD	109.109	108.723	106.915	108.987	107.447
LPG MB/CD	5.579	5.561	6.119	5.140	5.592
Sulfur tons/CD	104	106	114	185	185
SO_x emissions tons/CD	51	54	59	13	14
Processing MB/CD					
Reforming					
Total	47.8	45.8	46.5	47.8	46.6
For gasoline	44.3	42.3	43.0	44.3	43.1
Severity for gasoline	97.0	98.0	100.0	97.3	100.0
Catalytic cracking					
Untreated feed	56.6	60.5	62.2	–	–
Hydrotreated feed	–	–	–	62.2	68.1
Total	56.6	60.5	62.2	62.2	68.1
Conversion vol %	85.0	84.1	83.6	72.5	72.5
Hydrocracking					
High severity	5.4	4.2	1.4	4.1	0.1
Medium severity	3.8	5.1	8.0	6.1	9.4
Total	9.2	9.3	9.4	10.2	9.5
Isomerization of light naphtha					
Once through	0.2	3.2	0.1	0.1	2.2
Recycle	–	0.2	7.6	0.9	8.2
Total	0.2	3.4	7.7	1.0	10.4
Alkylation (product basis)	11.4	12.6	12.9	9.0	10.5
Hydrogen manufacture (MMSCF/CD)	24.2	22.7	20.7	24.2	17.9
Desulfurization					
Light naphtha (isom. feed)	0.2	3.4	7.7	1.0	10.4
Medium naphtha (ref. feed)	38.1	36.8	36.5	38.2	41.7
Cat cracker cycle oil	0.1	2.3	4.1	–	–
Cat cracker feed	–	–	–	62.5	68.5
Straight run distillate	19.7	13.7	12.9	31.4	17.4
Total	58.1	56.2	61.2	133.1	138.0

Table J-24. L.P. MODEL RESULTS – PROCESSING AND VARIABLE OUTPUTS

Large Midwest Cluster

	Scenario C			Scenario F	
	1977	1980	1985	1977	1985
Variable output					
Gasoline MB/CD	79.295	77.425	76.615	77.040	74.830
LPG MB/CD	3.121	3.899	4.063	3.415	4.236
Sulfur tons/CD	156	174	174	174	204
SO_x emissions tons/CD	61	71	73	13	14
Processing MB/CD					
Reforming					
Total	29.6	29.0	31.4	28.7	35.4
For gasoline	27.3	26.7	29.1	26.4	33.1
Severity for gasoline	96.5	100.0	100.0	100.0	100.0
Catalytic cracking					
Untreated feed	52.4	51.2	51.6	–	–
Hydrotreated feed	–	–	–	43.6	41.4
Total	52.4	51.2	51.6	43.6	41.4
Conversion vol %	66.6	75.6	72.4	72.5	72.5
Isomerization of light naphtha					
Once through	–	–	–	–	–
Recycle	–	–	7.0	–	7.5
Total	–	–	7.0	–	7.5
Alkylation (product basis)	10.6	12.8	12.0	11.4	10.9
Coking	14.3	13.9	14.0	–	–
Hydrogen manufacturing (MMSCF/CD)	–	–	–	14.2	15.4
Desulfurization					
Light naphtha (isom. feed)	–	–	7.0	–	7.5
Medium naphtha (ref. feed)	19.0	21.8	28.8	28.1	34.2
Medium coker naphtha (ref. feed)	2.5	2.5	2.5	2.3	2.3
Cat cracker cycle oil	13.4	9.2	10.7	–	–
Cat cracker feed	–	–	–	43.8	44.4
Straight run distillate	24.8	26.4	23.8	20.1	24.0
Total	59.7	59.9	72.8	94.3	112.4

Table J-25. L.P. MODEL RESULTS – PROCESSING AND VARIABLE OUTPUTS

Small Midcontinent Cluster

	Scenario C			Scenario F	
	1977	1980	1985	1977	1985
Variable output					
Gasoline MB/CD	36.732	35.472	34.016	35.541	32.775
LPG MB/CD	1.454	1.948	1.720	1.552	1.622
Sulfur tons/CD	14	16	19	21	27
SO_x emissions tons/CD	21	22	24	5	6
Processing MB/CD					
Reforming					
Total	14.7	14.0	14.2	14.6	14.9
For gasoline	10.4	9.7	9.9	10.3	10.6
Severity for gasoline	90.7	98.7	100	92.6	100.0
Catalytic cracking					
Untreated feed	17.7	17.9	18.9	–	–
Hydrotreated feed	–	–	–	18.1	19.2
Total	17.7	17.9	18.9	18.1	19.2
Conversion vol %	85	85	85	72.5	72.5
Isomerization of light naphtha					
Once through	–	–	1.4	–	1.9
Recycle	0.5	1.6	3.2	0.9	2.9
Total	0.5	1.6	4.6	0.9	4.8
Alkylation (product basis)	5.0	5.1	5.4	3.8	4.1
Coking	4.0	4.0	3.4	4.2	4.0
Desulfurization					
Light naphtha (isom. feed)	0.5	1.6	4.6	0.9	4.8
Medium naphtha (ref. feed)	14.0	13.3	13.5	13.9	14.2
Medium coker naphtha (ref. feed)	0.7	0.7	0.6	0.7	0.7
Cat cracker cycle oil	–	0.3	2.4	–	–
Cat cracker feed	–	–	–	18.2	19.3
Straight run distillate	1.2	0.8	–	–	–
Total	16.4	16.7	21.1	33.7	39.0

Table J-26. L.P. MODEL RES'JLTS – PROCESSING AND VARIABLE OUTPUTS

Louisiana Gulf Cluster

	Scenario C			Scenario F	
	1977	1980	1985	1977	1985
Variable output					
Gasoline MB/CD	118.346	116.110	112.488	116.558	113.770
LPG MB/CD	3.588	4.315	5.082	3.831	0.910
Sulfur tons/CD	58	61	60	71	71
SO_x emissions tons/CD	18	22	24	2	2
Processing MB/CD					
Reforming					
Total	30.3	31.7	35.5	37.2	39.1
For gasoline	30.3	31.7	35.5	37.2	39.1
Severity for gasoline	95.1	100.0	100.0	94.1	100.0
Catalytic cracking					
Untreated feed	68.3	59.4	53.9	–	–
Hydrotreated feed	17.9	26.3	26.3	68.1	76.0
Total	86.2	85.7	80.2	68.1	76.0
Conversion vol %	66.6	67.3	71.6	72.5	72.5
Hydrocracking					
High severity	–	–	–	5.1	–
Medium severity	8.8	8.8	8.8	3.5	8.8
Total	8.8	8.8	8.8	8.6	8.8
Isomerization of light naphtha					
Once through	–	–	4.7	–	6.5
Recycle	–	–	3.5	–	1.4
Total	–	–	8.2	–	7.9
Alkylation (product basis)	17.4	17.7	18.1	15.3	17.3
Coking	18.7	18.4	17.6	15.5	15.2
Hydrogen manufacture (MMSCF/CD)	23.9	21.8	17.5	18.8	16.2
Desulfurization					
Light naphtha (isom. feed)	–	–	8.2	–	7.9
Medium naphtha (ref. feed)	24.3	25.8	30.0	29.9	33.9
Medium coker naphtha (ref. feed)	3.2	3.2	3.1	2.9	2.8
Cat cracker feed	18.0	26.4	26.4	68.4	76.4
Straight run distillate	2.6	2.6	2.7	4.9	7.3
Total	48.1	58.0	70.4	106.1	128.3

Table J-27. L.P. MODEL RESULTS – PROCESSING AND VARIABLE OUTPUTS

Texas Gulf Cluster

	Scenario C			Scenario F	
	1977	1980	1985	1977	1985
Variable output					
Gasoline MB/CD	161.267	158.202	152.330	161.288	152.850
LPG MB/CD	7.303	7.516	8.014	6.919	7.667
Sulfur tons/CD	186	183	173	233	229
SO_x emissions tons/CD	74	90	92	25	35
Processing MB/CD					
Reforming					
Total	71.3	72.4	72.2	68.1	71.3
For gasoline	51.4	53.7	54.9	48.2	53.9
Severity for gasoline	99.3	99.7	100.0	98.2	100.0
Catalytic cracking					
Untreated feed	95.7	93.4	80.6	–	–
Hydrotreated feed	–	–	–	93.2	87.0
Total	95.7	93.4	80.6	93.2	87.0
Conversion vol %	68.2	68.9	79.2	72.5	72.5
Hydrocracking					
High severity	5.7	6.7	3.1	–	–
Medium severity	14.2	13.1	17.5	15.6	20.2
Total	19.9	19.8	20.6	15.6	20.2
Isomerization of light naphtha					
Once through	0.3	0.3	7.0	0.3	3.4
Recycle	2.3	6.1	8.3	3.8	12.6
Total	2.6	6.4	15.3	4.1	16.0
Alkylation (product basis)	18.7	18.5	19.1	18.1	16.8
Coking	18.8	18.4	16.9	15.2	14.6
Hydrogen Manufacture (MMSCF/CD)	41.0	41.5	40.7	29.3	38.0
Desulfurization					
Light naphtha (isom. feed)	2.6	6.4	15.3	4.1	16.0
Medium naphtha (ref. feed)	49.6	49.4	50.0	50.4	50.8
Medium coker naphtha (ref. feed)	3.2	3.1	3.0	2.8	2.7
Cat cracker cycle oil	23.2	22.0	11.9	–	–
Cat cracker feed	–	–	–	93.2	87.5
Straight run distillate	6.1	6.5	7.1	6.9	6.5
Total	84.7	87.4	87.3	157.8	163.4

Table J-28. L.P. MODEL RESULTS – PROCESSING AND VARIABLE OUTPUTS

West Coast Cluster

	Scenario C			Scenario F	
	1977	1980	1985	1977	1985
Variable output					
Gasoline MB/CD	69.634	70.398	70.290	69.609	69.730
LPG MB/CD	4.033	3.016	0.044	3.997	–
Sulfur tons/CD	198	178	171	221	187
SO_x emissions tons/CD	46	42	47	10	13
Processing MB/CD					
Reforming					
Total	38.0	36.2	38.0	38.8	36.9
For gasoline	22.2	20.4	22.2	23.0	21.1
Severity for gasoline	93.8	96.9	100.0	96.3	100.0
Catalytic cracking					
Untreated feed	28.1	37.3	38.2	–	–
Hydrotreated feed	–	–	–	26.7	33.9
Total	28.1	37.3	38.2	26.7	33.9
Conversion vol %	75.4	65.0	68.6	72.9	65.0
Hydrocracking					
High severity	–	–	–	–	–
Medium severity	27.4	27.4	27.4	27.4	27.4
Total	27.4	27.4	27.4	27.4	27.4
Isomerization of light naphtha					
Once through	–	–	2.5	–	1.9
Recycle	–	–	2.6	–	2.5
Total	–	–	5.1	–	4.4
Alkylation (product basis)	7.9	7.4	7.9	7.4	8.2
Coking	41.8	41.7	33.0	41.2	32.1
Hydrogen manufacture (MMSCF/CD)	51.5	51.7	51.5	51.5	51.5
Desulfurization					
Light naphtha (isom. feed)	–	–	5.1	–	4.4
Medium naphtha (ref. feed)	26.5	21.7	24.7	25.2	23.6
Medium coker naphtha (ref. feed)	7.7	7.7	6.1	7.7	6.0
Cat cracker cycle oil	3.6	7.8	7.8	–	–
Cat cracker feed	–	–	–	26.8	34.1
Straight run distillate	14.9	14.9	14.9	14.9	14.9
Total	52.7	52.1	58.6	74.6	83.0

Table J-29. L.P. MODEL RESULTS – PROCESSING AND VARIABLE OUTPUTS

Grassroots Refineries, 1985

	Scenario C			Scenario F		
	East of Rockies		West of Rockies	East of Rockies		West of Rockies
	–sour	–sweet		–sour	–sweet	
Variable output						
Gasoline MB/CD	91.2	91.2	88.5	93.4	93.4	90.8
LPG MB/CD	–	–	–	–	–	–
Sulfur tons/CD	311	12	141	365	18	207
SO_x emissions tons/CD	118	30	55	22	12	19
Processing MB/CD						
Reforming						
Total	44.4	53.2	42.4	47.8	52.1	43.6
For gasoline	44.4	53.2	42.4	47.8	52.1	43.6
Severity for gasoline	96.3	97.4	99.4	96.8	97.2	99.8
Catalytic cracking						
Untreated feed	40.3	35.8	30.5	–	–	–
Hydrotreated feed	–	–	–	42.5	35.2	27.2
Total	40.3	35.8	30.5	42.5	35.2	27.2
Conversion vol %	85.0	65.0	85.0	72.5	72.5	95.0
Hydrocracking						
High severity	4.8	4.9	6.9	10.3	5.3	14.8
Medium severity	17.8	9.2	32.4	16.2	4.7	42.2
Total	22.6	14.1	39.3	26.5	10.0	57.0
Isomerization of light naphtha						
Once through	–	0.2	4.9	–	1.7	2.1
Recycle	9.8	8.1	–	10.4	8.3	–
Total	9.8	8.3	4.9	10.4	10.0	2.1
Aklylation (product basis)	12.7	8.1	9.9	10.3	8.5	9.5
Coking	–	–	4.6	–	–	4.3
Hydrogen manufacture (MMSCF/CD)	52.8	26.8	78.0	64.9	19.4	111.3
Desulfurization						
Full range naphtha	53.5	58.2	37.4	54.2	59.4	38.8
Medium coker naphtha	–	–	0.8	–	–	0.8
Cat cracker feed	–	–	–	42.7	35.4	27.3
Straight run distillate	22.4	4.0	34.3	23.0	6.8	18.2
Vacuum bottoms	17.9	–	8.9	15.4	–	26.3
Total	93.8	62.2	81.4	135.3	101.6	111.4

Table J-30. L.P. MODEL RESULTS SUMMARY - GASOLINE BLENDING

Cluster: East Coast	**Year: 1977**	
Scenario	**C**	**F**
Premium pool		
Research octane clear	94.9	95.3
Motor octane clear	86.3	85.9
Volume MB/CD	16.37	16.35
Lead CC/USG	1.53	1.70
Sulfur PPM	304	46
Composition LV%		
Butanes	10.9	10.6
90 RON reformate	–	–
100 RON reformate	17.6	26.8
Cat cracker gasoline (untreated feed)	28.8	–
Cat cracker gasoline (desulfurized feed)	–	29.6
Alkylate	41.7	32.5
Light hydrocrackate	–	–
Isomerized light naphtha	1.0	0.5
Total	100.0	100.0
Regular pool		
Research octane clear	87.7	87.9
Motor octane clear	79.0	79.1
Volume MB/CD	58.92	58.85
Lead CC/USG	1.32	1.27
Sulfur PPM	438	84
Composition LV%		
BTX raffinate	2.4	2.4
Butanes	7.1	6.8
90 RON reformate	19.0	18.2
100 RON reformate	6.8	2.3
Cat cracker gasoline (untreated feed)	41.0	–
Cat cracker gasoline (desulfurized feed)	–	46.7
Alkylate	5.2	6.2
Straight run	18.5	17.4
Total	100.0	100.0

Table J-31. L.P. MODEL RESULTS - GASOLINE BLENDING

Cluster: East Coast	Year			
	1977		1985	
Scenarios	C	F	C	F
Lead-free pool				
Research octane clear	93.8	94.1	93.8	93.9
Motor octane clear	84.0	84.0	84.0	84.0
Volume MB/CD	33.82	33.79	106.91	107.45
Sulfur PPM	136	19	394	57
Composition LV%				
BTX raffinate	–	–	0.1	0.1
Butanes	6.9	6.0	7.3	7.2
90 RON reformate	2.7	–	–	–
100 RON reformate	50.3	57.4	30.9	30.9
Cat cracker gasoline (untreated feed)	13.2	–	34.6	–
Cat cracker gasoline (desulfurized feed)	–	11.2	–	36.8
Alkylate	4.6	–	12.1	9.8
Light hydrocrackate	7.7	8.0	2.1	2.0
Isomerized light naphtha	–	2.5	6.8	9.1
Natural gasoline	12.4	12.4	3.9	3.9
Straight run	2.2	2.5	2.2	0.2
Total	100.0	100.0	100.0	100.0
Total gasoline pool				
Research octane clear	90.7	90.9	93.8	93.9
Motor octane clear	81.6	81.7	84.0	84.0
Volume MB/CD	109.11	108.99	106.91	107.45
Lead CC/USG	0.94	0.94	–	–
Sulfur PPM	324	58	394	57

Table J-32. L.P. MODEL RESULTS – GASOLINE BLENDING

Cluster: Large Midwest	Year: 1977	
Scenarios	C	F
Premium pool		
Research octane clear	94.6	94.7
Motor octane clear	85.5	85.8
Volume MB/CD	3.96	3.85
Lead CC/USG	1.20	1.19
Sulfur PPM	394	48
Composition LV%		
Butanes	7.3	10.1
100 RON reformate	21.0	19.9
Cat cracker gasoline (untreated feed)	28.9	–
Cat cracker gasoline (desulfurized feed)	–	30.7
Alkylate	37.6	37.9
Straight run	5.2	1.4
Total	100.0	100.0
Regular pool		
Research octane clear	87.1	87.2
Motor octane clear	78.3	78.3
Volume MB/CD	49.97	48.54
Lead CC/USG	1.40	1.40
Sulfur PPM	863	237
Composition LV%		
BTX raffinate	1.8	1.9
Butanes	4.4	2.6
90 RON reformate	17.0	–
100 RON reformate	4.6	20.5
Cat cracker gasoline (untreated feed)	36.0	–
Cat cracker gasoline (desulfurized feed)	–	28.9
Alkylate	14.1	17.9
Coker Gasoline	2.8	6.2
Straight run	19.3	22.0
Total	100.0	100.0

Table J-33. L.P. MODEL RESULTS - GASOLINE BLENDING

Cluster: Large Midwest	Year			
	1977		1985	
Scenarios	C	F	C	F
Lead-free pool				
Research octane clear	94.9	95.6	94.0	94.0
Motor octane clear	84.0	84.0	84.0	84.0
Volume MB/CD	25.37	24.85	76.81	74.83
Sulfur PPM	572	60	748	148
Composition LV%				
Butanes	8.5	8.5	5.6	6.0
100 RON reformate	44.5	42.8	29.5	34.6
Cat cracker gasoline (untreated feed)	33.5	–	37.1	–
Cat cracker gasoline (desulfurized feed)	–	40.8	–	32.1
Alkylate	6.0	5.3	15.6	14.6
Isomerized light naphtha	–	–	8.7	9.6
Coker gasoline	–	–	1.8	1.7
Natural gasoline	2.5	2.6	0.6	0.5
Straight run	3.0	–	1.1	0.9
Total	100.0	100.0	100.0	100.0
Total gasoline pool				
Research octane clear	90.0	90.3	94.0	94.0
Motor octane clear	80.5	80.5	84.0	84.0
Volume MB/CD	79.30	77.04	76.81	74.83
Lead CC/USG	0.94	0.94	–	–
Sulfur PPM	738	171	746	148

Table J-34. L.P. MODEL RESULTS SUMMARY - GASOLINE BLENDING

Cluster: Small Midcontinent	Year: 1977	
Scenarios	**C**	**F**
Premium pool		
Research octane clear	90.5	91.0
Motor octane clear	82.3	82.9
Volume MB/CD	1.84	1.78
Lead CC/USG	2.25	2.12
Sulfur PPM	220	81
Composition LV%		
BTX raffinate	11.7	2.5
Butanes	9.6	6.2
Cat cracker gasoline (untreated feed)	35.6	–
Cat cracker gasoline (desulfurized feed)	–	45.3
Alkylate	31.7	31.5
Coker Gasoline	–	1.7
Straight run	11.4	12.8
Total	100.0	100.0
Regular pool		
Research octance clear	86.0	86.0
Motor octane clear	78.8	78.8
Volume MB/CD	23.14	22.39
Lead CC/USG	1.31	1.32
Sulfur PPM	136	61
Composition LV%		
BTX raffinates	8.3	10.5
Butanes	7.0	6.4
90 RON reformate	36.9	29.1
Cat cracker gasoline (untreated feed)	18.2	–
Cat cracker gasoline (desulfurized feed)	–	22.4
Alkylate	4.6	4.8
Coker gasoline	1.7	1.7
Natural gasoline	11.5	13.2
Straight run	11.8	11.9
Total	100.0	100.0

Table J-35. L.P. MODEL RESULTS - GASOLINE BLENDING

Cluster: Small Midcontinent	Year			
	1977		1985	
Scenarios	C	F	C	F
Lead-free pool				
Research octane clear	92.3	92.9	92.8	93.1
Motor octane clear	84.0	84.0	84.0	84.0
Volume MB/CD	11.75	11.37	34.02	32.78
Sulfur PPM	255	49	237	65
Composition LV%				
BTX raffinate	–	–	5.3	5.9
Butanes	8.0	8.3	7.5	7.1
100 RON reformate	4.2	17.1	22.4	24.6
Cat cracker gasoline (untreated feed)	45.9	–	32.1	–
Cat cracker gasoline (desulfurized feed)	–	39.6	–	33.4
Alkylate	28.6	19.2	15.8	12.6
Isomerized light naphtha	4.3	7.7	12.8	14.0
Coker gasoline	–	–	1.0	1.2
Natural gasoline	6.3	5.3	3.1	1.2
Straight run	2.7	2.8	–	–
Total	100.0	100.0	100.0	100.0
Total gasoline pool				
Research octane clear	88.3	88.4	92.8	93.1
Motor octane clear	80.6	80.6	84.0	84.0
Volume MB/CD	36.73	35.54	34.02	32.78
Lead CC/USG	0.94	0.94	–	–
Sulfur PPM	178	58	237	65

Table J-38. L.P. MODEL RESULTS SUMMARY - GASOLINE BLENDING

Cluster: Louisiana Gulf	Year: 1977	
Scenarios	**C**	**F**
Premium pool		
Research octane clear	96.77	97.2
Motor octane clear	87.1	87.4
Volume MB/CD	15.38	15.15
Lead CC/USG	0.74	0.60
Sulfur PPM	6	4
Composition LV%		
Butanes	10.2	10.2
100 RON reformate	47.2	53.7
Cat cracker gasoline (desulfurized feed)	9.4	3.2
Alkylate	33.2	32.9
Total	100.0	100.0
Regular pool		
Research octane clear	86.3	86.1
Motor octane clear	78.3	78.1
Volume MB/CD	66.28	66.27
Lead CC/USG	1.61	1.54
Sulfur PPM	332	90
Composition LV%		
Butanes	6.8	7.1
90 RON reformate	20.3	30.0
Cat cracker gasoline (untreated feed)	36.6	–
Cat cracker gasoline (desulfurized feed)	–	27.5
Alkylate	5.9	3.0
Light hydrocrackate	2.7	4.2
Coker gasoline	2.7	2.5
Natural gasoline	4.4	4.1
Straight run	20.6	21.6
Total	100.0	100.0

Table J-37. L.P. MODEL RESULTS - GASOLINE BLENDING

Cluster: Louisiana	Year			
	1977		1985	
Scenarios	C	F	C	F
Lead-free pool				
Research octane clear	94.0	94.0	93.6	93.6
Motor octane clear	84.0	84.0	84.0	84.0
Volume MB/CD	38.69	38.14	112.49	113.77
Sulfur PPM	230	76	217	84
Composition LV%				
Butanes	9.4	9.3	7.4	7.6
100 RON reformate	13.4	10.9	24.9	27.1
Cat cracker gasoline (untreated feed)	30.6	–	24.8	–
Cat cracker gasoline (desulfurized feed)	23.8	55.9	14.6	37.7
Alkylate	22.8	23.2	16.1	15.2
Light hydrocrackate	–	–	1.8	1.8
Isomerized light naphtha	–	–	6.9	6.7
Coker gasoline	–	–	1.5	1.4
Natural gasoline	–	0.7	2.0	2.0
Straight run	–	–	–	0.5
Total	100.0	100.0	100.0	100.0
Total gasoline pool				
Research octane clear	90.0	90.0	93.6	93.6
Motor octane clear	81.2	81.1	84.0	84.0
Volume MB/CD	118.35	116.56	112.49	113.77
Lead CC/USG	0.94	0.94	–	–
Sulfur PPM	258	74	217	84

Table J-38. L.P. MODEL RESULTS SUMMARY - GASOLINE BLENDING

Cluster: Texas Gulf	Year: 1977	
Scenarios:	**C**	**F**
Premium pool		
Research octane clear	91.5	91.4
Motor octane clear	83.5	83.1
Volume MB/CD	20.98	20.97
Lead CC/USG	2.02	2.24
Sulfur PPM	293	35
Composition LV%		
BTX raffinate	14.0	14.6
Butanes	8.8	8.8
Cat cracker gasoline (untreated feed)	37.3	–
Cat cracker gasoline (desulfurized feed)	–	42.1
Alkylate	25.7	21.3
Light hydrocrackate	13.5	12.4
Straight run	0.7	.8
Total	100.0	100.0
Regular Pool		
Research octane clear	90.1	90.2
Motor octane clear	80.8	80.9
Volume MB/CD	90.32	90.32
Lead CC/USG	1.20	1.16
Sulfur PPM	624	122
Composition LV%		
BTX raffinate	3.1	3.1
Butanes	6.6	7.5
90 RON reformate	3.5	5.6
95 RON reformate	–	6.4
100 RON reformate	9.1	–
Cat cracker gasoline (untreated feed)	47.9	–
Cat cracker gasoline (desulfurized feed)	–	47.6
Alkylate	14.7	15.1
Coker gasoline	1.4	1.8
Straight run	13.7	12.9
Total	100.0	100.0

Table J-39. L.P. MODEL RESULTS SUMMARY – GASOLINE BLENDING

	Year			
Cluster: Texas Gulf	1977		1985	
Scenarios	C	F	C	F
Lead free pool				
Research octane clear	93.2	93.3	93.2	93.5
Motor octane clear	84.0	84.0	84.0	84.0
Volume MB/CD	49.99	50.00	152.33	152.85
Sulfur PPM	44	8	309	65
Composition LV%				
BTX raffinate	–	–	2.1	2.1
Butanes	6.3	6.3	7.2	6.9
100 RON reformate	60.4	57.6	29.0	28.3
Cat cracker gasoline (untreated feed)	–	–	30.6	–
Cat cracker gasoline (desulfurized feed)	–	4.4	–	33.0
Alkylate	–	–	12.5	11.0
Light hydrocrackate	5.3	2.0	3.4	3.1
Isomerised light naphtha	5.0	7.8	9.5	10.0
Coker gasoline	1.1	–	1.1	1.0
Natural gasoline	21.9	21.9	4.6	4.6
Total	100.0	100.0	100.0	100.0
Total Gasoline Pool				
Research octane clear	91.2	91.3	93.2	93.5
Motor octane clear	82.1	82.1	84.0	84.0
Volume MB/CD	161.27	161.29	152.33	152.85
Lead CC/USG	0.94	0.94	–	–
Sulfur PPM	401	75	309	65

Table J-40. L.P. MODEL RESULTS SUMMARY - GASOLINE BLENDING

Cluster: West Coast	Year: 1977	
Scenarios	**C**	**F**
Premium pool		
Research octane clear	92.7	92.7
Motor octane clear	83.5	83.4
Volume MB/CD	15.32	15.31
Lead CC/USG	1.63	1.64
Sulfur PPM	1328	108
Composition LV%		
Butanes	6.4	6.2
100 RON reformate	0.6	–
Cat cracker gasoline (untreated feed)	50.1	–
Cat cracker gasoline (desulfurized feed)	–	57.7
Alkylate	24.4	24.4
Straight run	11.5	11.7
Total	100.0	100.0
Regular pool		
Research octane clear	85.9	85.9
Motor octane clear	78.1	78.1
Volume MB/CD	29.25	29.24
Lead CC/USG	1.38	1.38
Sulfur PPM	788	192
Composition LV%		
BTX raffinate	29.0	30.3
Butanes	7.7	7.8
90 RON reformate	22.4	22.0
Cat cracker gasoline (untreated feed)	20.0	–
Cat cracker gasoline (desulfurized feed)	–	19.1
Light hydrocrackate	8.7	8.6
Coker gasoline	4.1	3.4
Straight run	8.1	8.8
Total	100.0	100.0

Table J-41. L.P. MODEL RESULTS - GASOLINE BLENDING

Cluster: West Coast	Year			
	1977		1985	
Scenarios	C	F	C	F
Lead-free pool				
Research octane clear	92.1	92.3	93.3	93.2
Motor octane clear	84.0	84.0	84.0	84.0
Volume MB/CD	25.07	25.06	70.29	69.73
Sulfur PPM	171	17	779	47
Composition LV%				
BTX raffinate	.9	–	8.2	10.4
Butanes	7.2	7.4	7.3	7.5
90 RON reformate	22.0	23.9	–	–
100 RON reformate	24.1	26.2	26.2	25.0
Cat cracker gasoline (untreated feed)	5.7	–	29.9	–
Cat cracker gasoline (desulfurized feed)	–	4.2	–	28.1
Alkylate	16.4	14.6	11.3	11.7
Light hydrocrackate	15.1	15.2	9.0	9.1
Isomerized light naphtha	–	–	7.0	6.1
Natural gasoline	3.5	3.5	–	–
Straight run	5.1	5.0	1.1	2.1
Total	100.0	100.0	100.0	100.0
Total gasoline pool				
Research octane clear	89.6	89.7	93.3	93.2
Motor octane clear	81.4	81.4	84.0	84.0
Volume MB/CD	69.63	69.61	70.29	69.73
Lead CC/USG	0.94	0.94	–	–
Sulfur PPM	673	111	779	47

Table J-42. L.P. MODEL RESULTS SUMMARY – GASOLINE BLENDING

Cluster: Grassroots Refineries	East Coast sweet crude		East Coast sour crude	
Scenarios	C	F	C	F
Lead free pool				
Research octane clear	93.7	93.7	93.3	93.5
Motor octane clear	84.0	84.0	84.0	84.0
Volume MB/CD	91.20	93.40	91.20	93.40
Sulfur PPM	70	7	433	40
Composition LV%				
Butanes	6.7	6.9	6.1	5.5
90 RON reformate	14.7	15.0	14.9	13.0
100 RON reformate	33.7	31.5	22.1	25.6
Cat cracker gasoline (untreated feed)	20.4	–	26.6	–
Cat cracker gasoline (desulfurized feed)	–	21.7	–	26.4
Alkylate	8.9	9.1	13.9	11.0
Light hydrocrackate	4.3	3.1	6.2	7.9
Isomerised light naphtha	8.6	10.2	10.2	10.6
Straight run	2.7	2.5	–	–
Total	100.0	100.0	100.0	100.0

Table J-43. L.P. MODEL RESULTS SUMMARY - GASOLINE BLENDING

Cluster: Grassroots Refineries	**West Coast - Alaskan North Slope crude**	
Scenarios	C	F
Lead Free pool		
Research octane clear	93.2	93.3
Motor octane clear	84.0	84.0
Volume MB/CD	88.50	90.80
Sulfur PPM	331	31
Composition LV%		
Butanes	6.3	4.7
90 RON reformate	2.6	0.7
100 RON reformate	37.0	38.7
Cat cracker gasoline (untreated feed)	20.7	–
Cat cracker gasoline (desulfurized feed)	–	20.0
Alkylate	11.2	10.4
Light hydrocrackate	11.1	16.6
Isomerised light naphtha	5.4	2.2
Coker gasoline	0.5	0.5
Straight run	5.2	6.2
Total	100.0	100.0

Table J-44. L.P. MODEL RESULTS: – RESIDUAL FUEL OIL SULFUR LEVELS – 1977

Scenario	Cluster											
	East Coast		Large Midwest		Small Midcont.		Louisiana Gulf		Texas Gulf		West Coast	
	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S
C	13.37	2.0	7.26	0.48	0.23	0.33	5.19	0.60	15.321	1.43	31.89	0.12
F	13.37	1.3	7.26	0.94	0.23	0.23	5.19	0.11	15.321	1.29	31.89	0.43

Table J-45. L.P. MODEL RESULTS: – RESIDUAL FUEL OIL SULFUR LEVELS – 1985

	Cluster																	
	East Coast		Large Midwest		Small Midcont.		Louisiana Gulf		Texas Gulf		West Coast		East of Rockies Grassroots Sweet Crude		East of Rockies Grassroots Sour Crude		West of Rockies Grassroots	
Scenario	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S
C	13.234	1.8	7.21	.98	0.215	.33	5.032	.61	14.714	1.22	31.28	.69	32.9	0.41	32.9	1.97	45.5	1.63
F	13.234	1.03	7.21	1.13	0.215	.45	5.032	.30	14.714	1.42	31.28	.87	32.9	0.36	32.9	2.12	45.5	1.16

Table J-46. L.P. MODEL RESULTS: – REFINERY FUEL SULFUR LEVELS – 1977

Scenario	Cluster											
	East Coast		Large Midwest		Small Midcont.		Louisiana Gulf		Texas Gulf		West Coast	
	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S
C	14.46	0.6	10.09	1.5	4.80	1.3	15.68	0.2	26.62	0.9	16.63	0.7
F	14.96	0.3	10.97	0.5	4.68	0.5	16.21	<0.1	26.87	0.5	16.86	0.3

Table J-47. L.P. MODEL RESULTS: – REFINERY FUEL SULFUR LEVELS – 1985

	Cluster																	
	East Coast		Large Midwest		Small Midcont.		Louisiana Gulf		Texas Gulf		West Coast		East of Rockies Grassroots Sweet Crude		East of Rockies Grassroots Sour Crude		West of Rockies Grassroots	
Scenario	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S	MB/CD	WT % S
C	14.81	0.6	10.98	1.5	5.05	1.4	16.55	0.4	27.07	0.9	16.57	0.7	13.60	0.5	17.82	1.0	18.24	0.7
F	15.11	0.3	12.00	0.5	4.96	0.5	17.07	<0.1	27.20	0.5	16.93	0.3	13.90	0.4	18.63	0.4	20.82	0.3

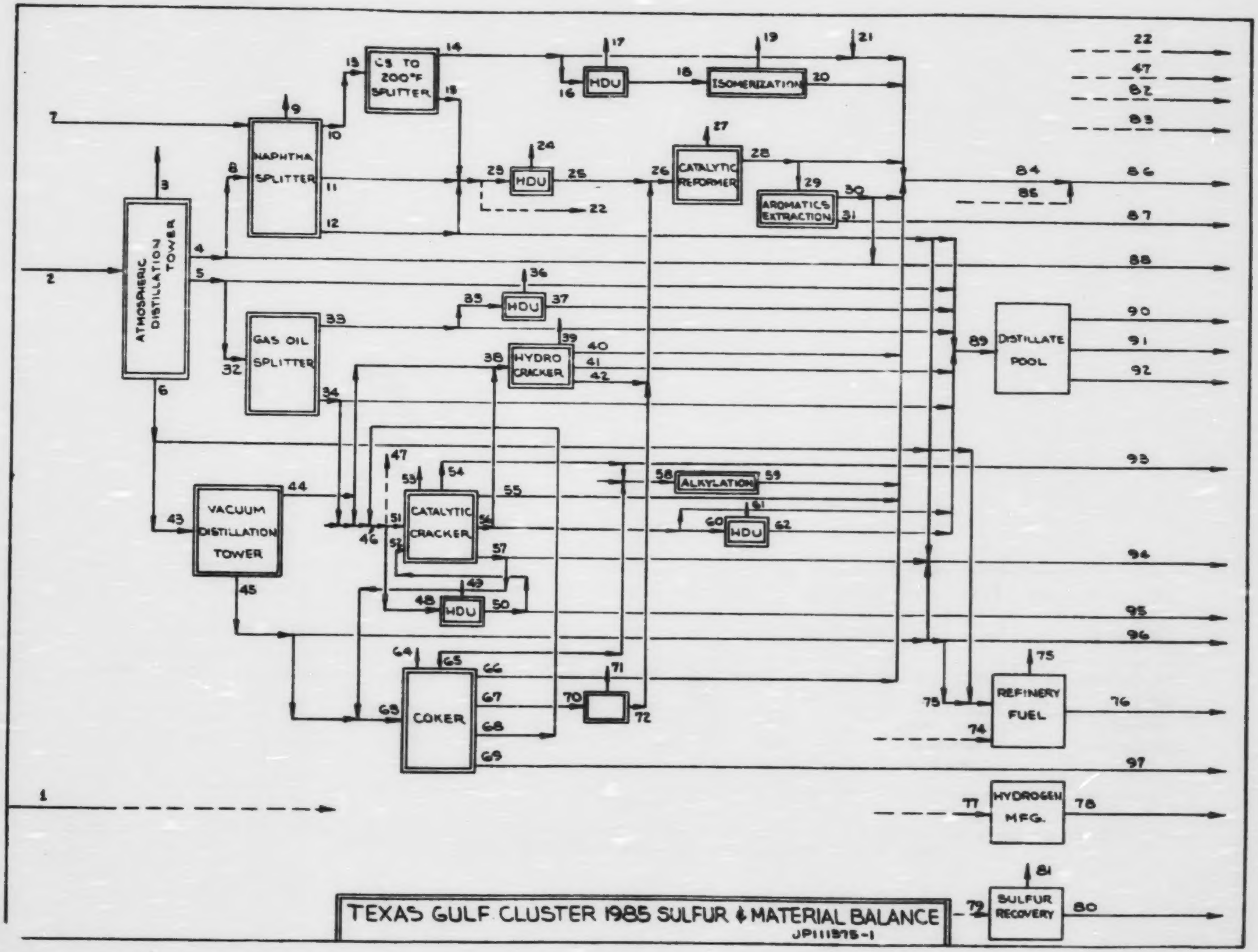
ATMOSPHERIC DISTILLATION TOWER
NAPHTHA SPLITTER
C3 TO 200°F SPLITTER
HDU
ISOMERIZATION
CATALYTIC REFORMER
AROMATICS EXTRACTION
GAS OIL SPLITTER
HYDRO CRACKER
VACUUM DISTILLATION TOWER
CATALYTIC CRACKER
ALKYLATION
COKER
DISTILLATE POOL
REFINERY FUEL
HYDROGEN MFG.
SULFUR RECOVERY
TEXAS GULF CLUSTER 1985 SULFUR & MATERIAL BALANCE
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Figure J-1

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Table J-48. SAMPLE CALCULATIONS FOR MASS AND SULFUR BALANCE

Texas Gulf 1985, Scenario B/C

Stream Values—Gas Oil 375-650°F

Gas oil/Crude charge	(1) Crude volume, MB/CD[a]	(2) Yield on crude, volume %[b]	(3) Specific gravity[c]	(4) Sulfur content, % weight[c]	(5) $\frac{(1) \times (2)}{100}$ Stream volume, MB/CD	(6) $(5) \times (3) \times 349.776$ Hydrocarbon weight, Mlbs/CD	(7) $\frac{(4) \times (6)}{100}$ Sulfur weight, Mlbs/CD	(8) $\frac{(7) \times 10^6}{(6)}$ Sulfur content, PPM
Louisiana	155.669	37.0	0.837	0.0649	57.598	16,862.535	10.944	649
West Texas Sour	135.964	27.5	0.8440	0.9146	37.390	11,037.937	100.953	9,146
Nigerian Forcados	12.480	38.7	0.874	0.1452	4.830	1,476.551	2.144	1,452
Arabian Light	17.405	28.4	0.8278	0.6849	4.943	1,430.873	9.800	6,849
Venezuelan Tia Juana	6.897	23.0	0.8473	0.4599	1.586	470.035	2.162	4,600
Total	328.415	N.A.	0.8409	0.4028	106.347	31,277.931	126.003	4,028[d]
Values for stream #5 gas oil 375-650°F					106.347	31,277.931	126.003	4,028

[a] Table J-20.

[b] Table H-1, sum of light and heavy gas oil yields.

[c] Tables H-13 and H-14.

[d] Average

Table J-49. SAMPLE CALCULATIONS FOR MASS AND SULFUR BALANCE

Texas Gulf 1985 B/C

Desulfurization of Light Gas Oil

Stream number	Stream name	(1) Process coefficients[a]	(2) (1)x7.080[b] Volume, MB/CD	(3) Specific gravity[c]	(4) Sulfur, % wgt.[c]	(5) (2)x(3)x349.776 Hydrocarbon weight, Mlbs/CD	(6) $\frac{(4) \times (5)}{100}$ Intake sulfur weight, Mlbs/CD	(7) Output sulfur distribution, percent[d]	(8) Output sulfur weight, Mlbs/CD	(9) Sulfur content, PPM
	Desulfurization of light gas oil									
35	Intake									
	Texas light gas oil	1.000	7.080	0.8251	0.5787	2,043.289	11.824			5,786.
	Normal purity hydrogen, MSCF/CD	0.190	1.345	N.A.	0.0	7.270[e]	0.000			0
	Output									
	C_3	0.001	0.007	0.508	negl.	1.244		0.0	0.000	0
	iC_4	0.001	0.007	0.563	negl.	1.378		0.0	0.000	0
	nC_4	0.001	0.007	0.584	negl.	1.430		0.0	0.000	0
	C_5 to 160°F	0.008	0.057	0.664	0.0288	19.937		0.0	0.000	0
	H_2S	N.A.	N.A.	N.A.	N.A.	12.440[f]		99.0	11.708	N.A.
36	Subtotal – light ends and H_2S		0.078			36.429		99.0	11.708	N.A.
37	Desulfurized light gas oil	0.990	7.009	0.815		1,998.038		1.0	0.117	58.

[a] Tables H-8 and H-10.

[b] Intake volume for undesulfurized light gas oil; MB/CD unless otherwise noted.

[c] Tables H-13 through H-15.

[d] Table H-12.

[e] Weight of hydrogen assumed to be 5.406 lb/MSCF.

[f] Sulfur in H_2S (column 8) times the weight ratio of H_2S/sulfur (1.0625).

Table J-50. SAMPLE CALCULATIONS FOR MASS AND SULFUR BALANCE

Texas Gulf 1985, Scenario B/C

Feed Sulfur Levels

Process unit / Feed stream	(1) % crude charge[a]	(2) Yield on crude, volume %[b]	(3) $\frac{(1) \times (2)}{100}$ Feed, MB/CD	(4) Specific gravity[c]	(5) Sulfur content, % weight[c]	(6) $\frac{(3) \times (4) \times 349.776 \times (5)}{100}$ Feed sulfur Mlbs/CD	(7) (d) Feed sulfur, lbs/bbl feed
Catalytic cracker							
Vacuum overhead feed							
– Louisiana	47.4	32.50	15.405	0.8974	0.3221	15.575	0.501
– West Texas Sour	41.4	29.60	12.254	0.9167	1.8513	72.740	2.342
– Nigerian Forcados	3.8	30.40	1.155	0.942	0.3125	1.189	0.038
– Arabian Light	5.3	29.50	1.584	0.9154	2.3215	11.624	0.374
– Venezuelan Tia Juana	2.1	32.80	0.689	0.922	1.6292	3.619	0.117
Total VOH feed	**100.0**	**N.A.**	**31.067**	**0.9081**	**1.0814**	**104.747**	**3.372**
Coker/Visbreaker							
Vacuum bottoms 1050°F+							
– Louisiana	47.4	5.60	2.654	0.9881	0.9207	8.445	0.906
– West Texas Sour	41.4	12.30	5.092	1.0187	3.0018	54.464	5.844
– Nigerian Forcados	3.8	8.50	0.323	0.998	0.6265	0.706	0.075
– Arabian Light	5.3	13.70	0.726	1.0195	4.5630	11.787	1.265
– Venezuelan Tia Juana	2.1	25.00	0.525	1.0236	2.8743	5.403	0.580
Total bottoms feed	**100.0**	**N.A.**	**9.320**	**1.0086**	**2.4651**	**80.805**	**8.670**
Hydrocracker							
Heavy gas oil 500-650°F feed							
– Louisiana	47.4	19.50	9.243	0.8504	0.0901	2.477	0.146
– West Texas Sour	41.4	14.11	5.842	0.8633	1.2187	21.499	1.271
– Nigerian Forcados	3.8	20.60	0.783	0.891	0.2015	0.492	0.029
– Arabian Light	5.3	15.01	0.780	0.8463	1.0807	2.495	0.148
– Venezuelan Tia Juana	2.1	12.70	0.267	0.865	0.6690	0.540	0.032
Total HGO feed	**100.0**	**N.A.**	**16.915**	**0.8588**	**0.5426**	**27.503**	**1.626**

[a] Table A-5. Volume for each crude as a percent of total crude charge.

[b] Table H-1.

[c] Tables H-13 and H-14.

[d] Crude-specific feed sulfur (column 6) divided by total feed to unit (column 3).

Table J-51. SAMPLE CALCULATIONS FOR MASS AND SULFUR BALANCE

Texas Gulf 1985, Scenario B/C

Stream Qualities – Cluster Specific Streams

Process unit / Output stream	Crude-specific sulfur distribution (1) Sulfur content, lbs/bbl feed[a] Louisiana VOH	Texas VOH	Nigerian VOH	Arabian VOH	Venezuela VOH	Total VOH	(2) Stream volume, LV fraction on feed[b]	(3) (1) ÷ (2) Sulfur content, lbs/bbl[c]	(4) Gaseous weight lbs/bbl feed[d]
Catalytic cracker									
65 conversion[e]									
Cat. naphtha	0.025	0.086	0.002	0.016	0.045	0.184	0.52	0.354	N.A.
Light cycle oil	0.090	0.726	0.007	0.090	0.005	0.918	0.27	3.400	N.A.
Heavy cycle oil	0.153	0.551	0.011	0.076	0.036	0.827	0.08	10.338	N.A.
H_2S	0.209	0.892	0.016	0.156	0.027	1.300	N.A.	N.A.	1.381
SO_x	0.024	0.077	0.002	0.036	0.004	0.143	N.A.	N.A.	0.286
Total Cat – 65 conversion	0.501	2.342	0.038	0.374	0.117	3.372	N.A.	N.A.	N.A.
85 conversion[f]									
Cat. naphtha	0.020	0.073	0.002	0.013	0.053	0.161	0.60	0.268	N.A.
Light cycle oil	0.069	0.628	0.005	0.074	0.004	0.780	0.10	7.800	N.A.
Heavy cycle oil	0.132	0.452	0.010	0.060	0.031	0.685	0.05	13.700	N.A.
H_2S	0.244	1.066	0.018	0.182	0.022	1.522	N.A.	N.A.	1.617
SO_x	0.036	0.133	0.003	0.045	0.007	0.224	N.A.	N.A.	0.448
Total Cat – 85 conversion	0.501	2.342	0.038	0.374	0.117	3.372	N.A.	N.A.	N.A.

[a] Crude-specific feed sulfur (Table J-50, column 7) times output stream sulfur as a percent of crude-specific feed sulfur (Table H-18, case 2).

[b] Table H-4.

[c] Sulfur content, PPM, calculated by methodology shown on Table J-48. Specific gravities of streams found on Table H-16.

[d] Gaseous sulfur content (column 1) times weight ratio of gaseous stream/sulfur.

H_2S/sulfur weight ratio = 1.0625; SO_x/sulfur content = 2.000.

[e] Low conversion, untreated feed.

[f] High conversion, untreated feed.

Table J-52. SAMPLE CALCULATIONS FOR MASS AND SULFUR BALANCE

Texas Gulf 1985, Scenario B/C

Stream Qualities – Cluster-Specific Streams

Process/Stream[a]	(1) Total feed sulfur lbs/bbl feed[b]	(2) Stream sulfur, % feed[c]	(3) $\frac{(1) \times (2)}{100}$ Output sulfur, lbs/bbl feed	(4) Stream volume, LV fraction on feed[d]	(5) (3) ÷ (4) Sulfur content, lbs/bbl[e]	(6) (f) Gaseous weight, lbs/bbl feed
Catalytic cracker						
72.5 conversion[g]						
Cat. naphtha	0.337	3.5	0.012	0.58	0.021	N.A.
Light cycle oil	0.337	34.5	0.116	0.212	0.547	N.A.
Heavy cycle oil	0.337	33.5	0.113	0.063	1.794	N.A.
H_2S	0.337	20.0	0.067	N.A.	N.A.	0.071
SO_x	0.337	8.5	0.029	N.A.	N.A.	0.058
Total cat – 72.5 conversion	0.337	100.0	0.337	N.A.	N.A.	0.129
Coker						
Vacuum bottoms feed						
Light coker naphtha	8.670	1.2	0.104	0.105	0.990	N.A.
Medium coker naphtha	8.670	3.4	0.295	0.187	1.578	N.A.
Coker gas oil	8.670	30.3	2.627	0.413	6.361	N.A.
Coke	8.670	30.7	2.662	0.258	10.318	N.A.
H_2S	8.670	34.4	2.982	N.A.	N.A.	3.168
Total coker – bottoms feed	8.670	100.0	8.670	N.A.	N.A.	3.168
Heavy cycle oil feed[h]						
Light coker naphtha	10.338	1.2	0.124	0.080	1.550	N.A.
Medium coker naphtha	10.338	3.4	0.351	0.142	2.472	N.A.
Coker gas oil	10.338	30.3	3.133	0.5932	5.282	N.A.
Coke	10.338	30.7	3.174	0.1283	24.739	N.A.
H_2S	10.338	34.4	3.556	N.A.	N.A.	3.778
Total coker – cycle oil feed	10.338	100.0	10.338	N.A.	N.A.	3.778
Hydrocracker						
Heavy gas oil feed						
Output streams	1.626	0.0[i]	0.000[i]	N.A.	0.000[i]	N.A.
H_2S	1.626	100.0	1.626	N.A.	N.A.	1.728
Total hydrocracker – HGO	1.626	100.0	1.626	N.A.	0.000	1.728
Vacuum overhead feed						
Output streams	3.372	0.0[i]	0.000[i]	N.A.	0.000[i]	N.A.
H_2S	3.372	100.0	3.372	N.A.	N.A.	3.583
Total hydrocracker – VOH	3.372	100.0	3.372	N.A.	0.000	3.583
Light cycle oil feed[h]						
Output streams	3.400	0.0[i]	0.000[i]	N.A.	0.000[i]	N.A.
H_2S	3.400	100.0	3.400	N.A.	N.A.	3.613
Total hydrocracker – LCO	3.400	100.0	3.400	N.A.	0.000	3.613

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[a] Other cat cracker conversions on Table J-51. Visbreaking not used in Texas Gulf cluster.

[b] Table J-50.(column 7).

[c] Tables H-17 and H-18 (case 2).

[d] Table H-4 through H-7.

[e] Sulfur content, PPM, calculated by methodology shown on Table J-48. Specific gravities of streams found on Table H-16.

[f] Gaseous output sulfur (column 3) times weight ratio of gaseous stream/sulfur.
H_2S/sulfur weight ratio = 1.0625, SO_x/sulfur weight ratio = 2.000.

[g] Low conversion, hydrotreated feed (sulfur level is 10% of untreated feed). High (95) conversion, hydrotreated feed not used in Texas Gulf.

[h] 85 conversion catalytic cracker output (Table J-51).

[i] Negligible sulfur content (approximately 1 PPM).

Table J-53. SPECIFIC GRAVITIES AND DENSITIES FOR MISCELLANEOUS STREAMS

Stream	Spgr	Lbs/MSCF
Refinery gas (FOE)	0.9714	
BTX	0.872	
Mixed olefins	0.550	
Coke	1.100	
Hydrogen		5.405

Table J-54. MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake				Process output			
		Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
1	Purchased butanes	2.926	586.637	0.293	499.				
	Atmospheric distillation tower								
2	Intake								
	Crude charge	328.415	97,946.709	751.830	7,676.				
	Output								
3	Light ends					5.627	1,124.336	0.000	0
4	Full range naphtha					83.708	22,068.196	16.435	744.
5	Gas oil 375 to 650°F					106.347	31,277.931	126.003	4,028.
6	Bottoms 650° F+					132.642	43,221.001	609.392	14,099.
	Naphtha splitter								
	Intake								
7	Purchased natural gasoline	11.200	2,634.901	0.076	29.				
8	Full range naphtha	80.070	21,134.135	16.162	764.				
	Output								
9	Light ends					1.376	277.400	0.000	0
10	C_5 to 200°F					33.952	8,156.236	1.258	154.
11	200 to 340°F					45.375	12,328.216	10.062	816.
12	340 to 375°F					10.564	2,976.906	4.921	1,653.
	C_5 to 200° splitter								
13	Intake								
	C_5 to 200°F	33.952	8,156.225	1.258	154.				
	Output								
14	C_5 to 160°F					21.864	5,037.023	0.551	109.
15	160 to 200°F					12.088	3,080.919	0.701	227.
	Desulfurization of isomerization feed								
16	Intake								
	C_5 to 160°F	14.900	3,601.199	0.577	160.				
	Normal purity hydrogen	1.564	8.454	0.000	0				
	Output								
17	H_2S					N.A.	0.582	0.548	N.A.
18	C_5 to 160°F desulfurized					14.900	3,467.882	0.004	1.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-54.(continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake				Process output			
		Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Isomerization								
18	Intake								
	C_5 to 160°F desulfurized	15.230	3,545.238	0.004	1.				
	Output								
19	Light ends					0.482	163.902	0.000	0
20	Isomerate					14.493	3,265.166	0.003	1.
	Naphtha product from desulfurization								
21	C_5 to 160° from LGO desulfurization	0.317	75.285	0.016	212.	0.317	75.285	0.016	212.
	Reformer feed to transfer								
22	SR naphtha	4.650	1,239.361	0.115	92.	4.650	1,239.361	0.115	92.
	Desulfurization SR naphtha								
23	Intake								
	SR naphtha 160 to 375°F	61.683	16,626.918	13.534	813.				
	Normal purity hydrogen	6.478	35.016	0.000	0				
	Output								
24	H_2S					N.A.	14.435	13.585	N.A.
25	Desulfurized SR naphtha					61.683	16,626.918	0.019	1.
	Catalytic reforming								
26	Intake								
	SR naphtha (desulf. and undesulf.)	62.463	16,839.077	0.018	1.				
	Heavy hydrocrackate	6.793	1,806.542	0.007	4.				
	Medium coker naphtha	2.980	793.215	0.003	4.				
	Total intake	72.236	19,437.834	0.028	1.				
	Output								
27	Light ends					14.952	2,837.415	0.001	0.
28	Reformate					57.614	16,290.655	0.019	1.
	Aromatics extraction								
29	Intake								
	100 sev. reformate	13.470	3,782.611	0.004	1.				
	Output								
30	Raffinate					7.795	1,980.780	0.002	1.
31	BTX					5.675	1,731.480	0.000	0

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-54.(continued). **MASS AND SULFUR BALANCE**

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake				Process output			
		Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Gas oil splitter								
32	Intake								
	Gas oil 375 to 650°F	59.210	17,415.680	115.823	6,650.				
	Output								
33	Light gas oil 375 to 500°F					28.484	8,198.376	33.001	4,029.
34	Heavy gas oil 500 to 650°F					30.726	9,226.711	82.872	8,981.
	Desulfurize light gas oil								
35	Intake								
	Light gas oil 375 to 500°F	7.080	2,043.289	11.824	5,786.				
	Normal purity hydrogen	1.354	7.270	0.000	0				
	Output								
36	Light ends and H_2S					0.078	36.429	11.708	N.A.
37	Desulfurized light gas oil					7.009	1,998.038	0.117	58.
21	C_5 to 160° naphtha					.317	75.285	0.016	212.
	Hydrocracker								
38	Intake								
	Hydrocarbon feed	20.646	6,621.258	68.634	10,366.				
	High purity hydrogen	40.725	220.134	0.000	0				
	Total intake		6,841.392	68.634	10,032				
	Output								
39	Light ends and H_2S					4.182	890.978	67.581	N.A.
40	Light hydrocrackate					5.235	1,244.852	0.002	0
41	Hydrocracked jet fuel					11.050	3,138.540	0.011	3.
42	Heavy hydrocrackate					6.793	1,805.542	0.007	3.
	Vacuum distillation tower								
43	Intake								
	Bottoms 650°+F	116.440	38,059.579	587.862	15,445.				
	Output								
44	Vacuum Overhead					88.206	28,069.637	330.035	11,757.
45	Bottoms 1050° + F					28.234	9,988.223	257.865	25,816.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-54.(continued). **MASS AND SULFUR BALANCE**

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake				Process output			
		Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
46	Gas oil feedstock	98.050	30,586.753	411.399	13,449.				
47	Cat. feed to transfer	1.880	586.469	7.888	13,449.				
	Desulfurization for lubes								
48	Input								
	Hydrocarbon feed	16.580	5,178.623	59.650	11,519.				
	Hydrogen		26.919	0.000	0				
	Output								
49	Light ends and H_2S					0.264	114.998	50.378	N.A.
50	Desulfurized product for lubes					16.517	5,087.446	5.566	1,092.
	Catalytic cracker								
51	Input	80.633	25,150.105	274.295	10,906				
	Output								
53	Light ends, H_2S and sulfur in SO_x					12.997	3,149.495	137.729	N.A.
54	Mixed olefins					13.163	2,532.256	0.000	0
55	Cat. naphtha					46.525	12,301.881	10.430	847.
56	Light cycle oil					12.072	3,781.617	55.696	14,728.
57	Heavy cycle oil					4.731	1,563.756	64.804	41,441.
	Alkylation								
58	Input								
	Isobutane	12.821	2,524.683	0.001	0				
	Mixed olefins	10.780	2,073.821	0.000	0				
59	Output								
	Alkylate					19.080	4,684.612	0.019	0

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-54.(continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake				Process output			
		Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Desulfurization of light cycle oil								
60	Input								
	Light cycle oil	11.850	3,712.124	55.573	14,970.				
	Normal purity hydrogen		14.092	0.000	0				
	Output								
61	Light ends and H_2S					0.131	92.790	47.295	N.A.
62	Desulfurized light cycle oil					11.731	3,675.003	8.251	2,245.
	Coker								
63	Input	16.900	5,901.070	151.456	25,666.				
	Output								
64	Light ends and H_2S					4.416	1,122.711	46.365	N.A.
65	Mixed olefins					0.616	118.503	0.000	0
66	Light coker naphtha					1.657	392.375	1.633	4,162.
67	Medium coker naphtha					2.950	789.357	4.628	5,863.
68	Coker gas oil					7.820	2,308.549	49.507	21,445.
69	Coke					3.754	1,444.365	38.549	26,689.
	Desulfurization of Coker Naphtha								
70	Input								
	Medium coker naphtha	2.95	789.357	4.629	5,864.				
	Normal purity hydrogen		9.588	0.000	0				
	Output								
71	H_2S					N.A.	4.915	4.626	N.A.
72	Desulfurized medium coker naphtha					2.980	793.215	0.003	4.
	Refinery Fuel System								
	Input (FOE)								
73	Bottoms 650° +F	18.420	5,550.302	62.467	11,255.				
74	Gases C_4 and lighter	10.670	1,177.801	0.000	0				
	Output								
75	Sulfur in SOX					N.A.	62.467	62.467	N.A.
76	Refinery fuel (FOE)					27.090	6,728.100	62.467	9,284.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-54.(continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake				Process intake			
		Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Hydrogen manufacturing								
77	Input								
	Methane/ethane (FOE)	1.702	544.251	0.000	0				
78	Output								
	High purity hydrogen					40.72	220.108	0.000	0
	Sulfur recovery								
79	Input								
	H_2S	N.A.	387.384	364.597	N.A.				
	Output								
80	Elemental sulfur					N.A.	346.342	346.342	N.A.
81	Sulfur in SO_X					N.A.	18.255	18.255	N.A.
	Blending of refinery gas								
82	Composition								
	Methane/ethane	0.788	267.578	0.000	0				
	Blend totals					0.788	267.578	0.000	0
	Blending of LPG								
83	Composition								
	Propane	8.014	1,423.938	0.000	0				
	Blend totals					8.014	1,423.938	0.000	0
	Blending of unleaded gasoline								
	Composition								
84	Liquid								
	Isomerate	14.493	3,265.166	0.003	1.				
	Reformate	44.144	12,508.044	0.015	1.				
	Light hydrocrackate	5.235	1,244.852	0.002	0				
	Cat. naphtha	46.525	12,301.881	10.430	847.				
	Light coker naphtha	1.657	392.375	1.633	4,164.				
	Raffinate	3.175	863.946	0.001	1.				
	Natural gasoline	6.981	1,562.352	0.016	10.				
	Alkylates	19.080	4,664.612	0.019	4.				
85	Gaseous								
	Butanes	11.040	2,253.092	0.000	0				
86	Blend totals					152.330	39,056.31	12.118	309.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-54.(continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake: Volume, MB/CD[a]	Process intake: Hydrocarbons weight, Mlbs/CD	Process intake: Sulfur weight, Mlbs/CD	Process intake: Sulfur content, PPM	Process output: Volume, MB/CD[a]	Process output: Hydrocarbons weight, Mlbs/CD	Process output: Sulfur weight, Mlbs/CD	Process output: Sulfur content, PPM
	Blending of BTX								
87	Composition								
	BTX	5.677	1,731.480	0.000	0				
	Blend totals					5.677	1,731.480	0.000	0
	Blending of naphtha								
88	Composition								
	SR naphtha	3.638	932.853	0.273	292.				
	Raffinate	4.620	1,116.834	0.000	0				
	Blend totals					8.258	2,049.680	0.273	133.
	Blending of distillates								
89	Composition								
	Heavy naphtha 340 to 375°F	0.814	228.249	0.048	210.				
	Gas oil 375 to 650°F	47.138	13,862.697	10.183	734.				
	Light gas oil	21.404	6,155.087	21.177	3,440.				
	Desulfurized light gas oil	7.009	1,998.095	0.117	58.				
	Hydrocracked jet fuel	11.050	3,138.540	0.011	3.				
	Heavy gas oil	8.055	2,396.789	2.159	900.				
	Light cycle oil	0.222	68.906	0.120	1,741.				
	Desulfurized light cycle oil	11.731	3,675.003	8.251	2,245.				
	Blend totals								
90	Jet fuel					22.043	6,269.908	0.827	131.
91	Kerosene					7.226	2,067.487	2.168	1,048.
92	Distillate fuel oil					78.141	23,177.402	39.071	1,685.
	Blending of olefins								
93	Composition								
	Mixed olefins	2.980	573.283	0.000	0				
	Light ends	0.745	135.790	0.000	0				
	Blend totals					3.725	709.073	0.000	0

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-54.(continued). **MASS AND SULFUR BALANCE**

Texas Gulf Cluster 1985, Scenario B/C

Stream number	Process/Stream name	Process intake				Process output			
		Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume, MB/CD[a]	Hydrocarbons weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Blending of residual fuel oil								
94	Composition								
	Heavy naphtha 340 to 375°F	0.111	31.681	0.003	95.				
	Bottoms 650° + F	10.379	3,307.285	13.834	4,183.				
	Bottoms 1050° + F	4.224	1,505.082	45.180	30,018.				
	Blend totals					14.714	4,844.048	59.017	12,183.
	Blending of lubes								
95	Composition								
	Desulfurized feed	15.474	4,766.188	5.205	1,092.				
	Blend totals					15.474	4,766.188	5.206	1,092.
	Blending of asphalt								
96	Composition								
	Bottoms 1050° + F	1.314	468.560	21.334	45,531.				
	Blend totals					1.314	468.560	21.334	45,531.
	Blending of coke								
97	Composition								
	Coke	3.754	1,444.365	38.550	26,689.				
	Blend totals					3.754	1,444.365	38.550	26,689.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-55. MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario F

Stream number	Process/stream name	Process intake				Process output			
		Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
1	Purchased butanes	2.926	586.637	0.293	499.				
	Atmospheric Distillation								
2	Intake								
	Crude charge	328.415	97,946.709	751.830	7,676.				
	Output								
3	Light ends					5.627	1,124.336	0.000	0.
4	Full range naphtha					83.708	22,088.196	16.435	744.
5	Gas oil 375 to 650°F					106.347	31,277.931	126.003	4,028.
6	Bottoms 650°+F					132.642	43,221.001	609.392	14,099.
	Naphtha splitter								
	Intake								
7	Purchased natural gasoline	11.200	2,634.901	.076	29.				
8	Full range naphtha	80.050	21,233.742	16.238	764.				
	Output								
9	Light ends					1.377	277.401	0.000	0.
10	C_5 to 200°F					33.946	8,154.832	1.258	154.
11	200 to 340°F					45.364	12,325.287	10.061	816.
12	340 to 375°F					10.561	2,976.073	4.921	1,653.
	C_5 to 200°F splitter								
13	Intake								
	C_5 to 200°F	33.045	8,104.170	1.257	155.				
	Output								
14	C_5 to 160°F					21.684	4,938.723	0.549	111.
15	160 to 200°F					12.046	3,070.121	0.700	228.
	Desulfurization of isomerization feed								
16	Intake								
	C_5 to 160°F	15.660	3,708.665	0.619	166.				
	Normal purity hydrogen	1.644	8.886	0.000	0.				
	Output								
17	H_2S					N.A.	0.622	0.586	N.A.
18	C_5 to 160°F desulfurized					15.680	3,842.508	0.005	1.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-55. (continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario F

		Process intake				Process output			
Stream number	Process/stream name	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Isomerization								
18	Intake								
	C_5 to 160°F desulfurized	16.000	4,074.715	0.005	1.				
	Output								
19	Light ends					.535	177.009	0.000	0.
20	Isomerate					15.187	3,417.954	0.004	1.
21	**Naphtha product from desulfurization**	0.052	12.031	0.003	249.	0.052	12.031	0.003	249.
	Reformer feed to transfer								
22	SR naphtha	4.650	1,239.361	0.115	92.	4.650	1,239.361	0.115	92.
	Desulfurization of SR naphtha								
23	Intake								
	SR naphtha 160 to 375°F	62.410	16,836.012	13.552	804.				
	Normal purity hydrogen	6.582	35.578	0.000	0.				
	Output								
24	H_2S					N.A.	13.800	12.988	N.A.
25	Desulfurized SR naphtha					62.410	16,836.012	.018	1.
	Catalytic reforming								
26	Intake								
	SR naphtha (desulf. and undesulf.)	61.403	17,040.361	0.018	1.				
	Heavy hydrocrackate	5.360	1,424.657	0.005	4.				
	Medium coker naphtha	2.750	731.993	0.003	4.				
	Total intake	69.513	19,197.011	0.026	1.				
	Output								
27	Light ends					16.258	3,127.796	0.001	0.
28	Reformate					56.839	16,451.301	0.019	1.
	Aromatics extraction								
29	Intake								
	100 sev. reformate	13.504	3,793.598	0.004	1.				
	Output								
30	Raffinate					7.827	1,990.253	0.002	1.
31	BTX					5.677	1,731.480	0.000	0.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-55. (continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario F

Stream number	Process/stream name	Process intake				Process output			
		Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Gas oil splitter								
32	Intake								
	Gas oil 375 to 650°F	59.540	17,583.471	117.135	6,661.				
	Output								
33	Light gas oil 375 to 500°F					28.614	8,260.803	33.312	4,032.
34	Heavy gas oil 500 to 650°F					30.926	9,322.812	83.848	8,993.
	Desulfurization of light gas oil								
35	Intake								
	Light gas oil 375 to 500°F	6.470	1,867.243	10.805	5,788.				
	Normal purity hydrogen	1.229	6.645	0.000	0.				
	Output								
36	Light ends and H_2S					N.A.	33.806	10.703	N.A.
37	Desulfurized light gas oil					6.405	1,825.942	0.107	58.
	Hydrocracker								
38	Intake								
	Hydrocarbon feed	20.200	6,392.135	67.956	10,631.				
	High purity hydrogen	37.996	205.384	0.000	0.				
	Total intake		6,597.519	67.956	10,300.				
	Output								
39	Light ends and H_2S					N.A.	1,012.792	67.451	N.A.
40	Light hydrocrackate					4.889	1,110.510	0.001	0.
41	Hydrocracked jet fuel					12.733	3,616.402	0.003	0.
42	Heavy hydrocrackate					5.356	1,461.252	0.006	3.
	Vacuum distillation tower								
43	Input								
	Bottoms 650° +F	117.240	37,677.285	588.926	15,630.				
	Output								
44	Vacuum overhead					88.890	28,314.306	330.823	11,683.
45	Bottoms 1050° +F					28.351	10,028.877	258.239	25,749.
46	Gas oil feedstock	104.890	32,739.498	389.387	11,893.				

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-55. (continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario F

Stream number	Process/stream name	Process intake				Process output			
		Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
47	**Cat. feed to transfer**	1.880	586.790	6.979	11,893.				
	Desulfurization of cat. feed								
48	Input								
	Undesulfurized cat. feed	103.023	32,152.708	382.408	11,893.				
	Hydrogen	30.907	167.065	0.000	0.				
	Output								
49	Light ends and H_2S					N.A.	880.728	325.018	N.A.
50	Desulfurized cat. feed					102.508	31,573.739	34.545	1,094.
	Catalytic cracker								
52	Input	87.004	26,807.427	29.330	1,094.				
	Output								
53	Light ends, H_2S and sulfur in SO_x					N.A.	2,266.940	8.422	N.A.
54	Mixed olefins					11.924	2,293.900	0.000	0.
55	Cat. naphtha					50.497	13,335.298	1.010	75.
56	Light cycle oil					18.451	5,779.941	10.111	1,749.
57	Heavy cycle oil					5.483	1,811.956	9.826	5,422.
	Alkylation								
58	Input								
	Isobutane	11.312	2,310.752	0.001	1.				
	Mixed olefins	5.522	1,062.209	0.000	0.				
59	Output								
	Alkylate					16.834	4,115.795	0.017	4.
	Desulfurization of light cycle oil								
60	Input	0.000	0.000	0.000	0.				
61/62	Output					0.000	0.000	0.000	0.
	Coker								
63	Input	14.550	5,185.064	132.090	25,475.				
	Output								
64	Light ends and H_2S					N.A.	1,204.282	43.193	N.A.
65	Mixed olefins					.567	109.156	0.000	0.
66	Light coker naphtha					1.528	381.757	1.506	4,163.
67	Medium coker naphtha					2.721	728.055	4.289	5,863.
68	Coker gas oil					6.009	1,774.011	38.045	21,445.
69	Coke					3.754	1,444.362	38.550	26,689.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-55. (continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario F

Stream number	Process/stream name	Process intake				Process output			
		Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Desulfurization of Coker naphtha								
70	Input								
	Medium coker naphtha	2.721	728.055	4.269	5,863.				
	Normal purity hydrogen	1.633	8.824	0.000	0.				
	Output								
71	H_2S					N.A.	4.533	4.266	N.A.
72	Desulfurized medium coker naphtha					2.748	731.488	0.003	4.
	Refinery fuel system								
	Input (FOE)								
73	Bottoms 650° + F	18.590	5,950.004	35.131	5,902.				
74	Gases (C_4 and lighter)	8.610	1,029.097	0.000	0.				
	Output								
75	Sulfur in SO_x					N.A.	35.121	35.121	N.A.
76	Refinery fuel (FOE)					27.200	6,979.101	35.121	5,032.
	Hydrogen manufacturing								
77	Input								
	Methane/ethane (FOE)	1.588	539.599	0.000	0.				
78	Output								
	High purity hydrogen					37.996	205.383	0.000	0.
	Sulfur recovery								
79	Input								
	H_2S	N.A.	466.936	458.293	N.A.				
	Output								
80	Elemental sulfur					N.A.	458.077	458.077	N.A.
81	Sulfur in SO_x					N.A..	0.504	0.252	N.A.
	Blending of refinery gas								
82	Composition (FOE)								
	Normal butane	0.790	256.147	0.000	0.	0.790	256.147	0.000	0.
	Blend totals (FOE)								
	Blending of LPG								
83	Composition								
	Propane	7.670	1,362.853	0.000	0.	7.670	1,362.853	0.000	0.
	Blend totals								

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-55. (continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario F

Stream number	Process/stream name	Process intake				Process output			
		Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Blending of unleaded gasoline								
	Composition								
84	Liquid								
	Isomerate	15.187	3,417.954	0.004	1.				
	Reformate	43.330	12,657.703	0.015	1.				
	Light hydrocrackate	4.669	1,110.510	0.001	0.				
	Cat. naphtha	50.497	13,335.298	1.010	75.				
	Light coker naphtha	1.528	361.757	1.506	4,163.				
	Raffinate	3.224	876.314	0.001	1.				
	Natural gasoline	6.983	1,572.596	0.016	10.				
	Alkylate	16.834	4,115.795	0.017	4.				
85	Gaseous								
	Butanes	10.598	2,153.220	0.003	1.				
86	Blend totals					152.850	39,601.147	2.573	65.
	Blending of BTX								
87	Composition								
	BTX	5.677	1,731.480	0.000	0.				
	Blend totals					5.677	1,731.480	0.000	0.
	Blending of naphtha								
88	Composition								
	SR naphtha	3.660	939.012	0.275	2.				
	Raffinate	4.600	1,113.939	0.001	1.				
	Blend totals					8.260	2,052.951	0.276	134.
	Blending of distillates								
89	Composition								
	Heavy naphtha 340 to 375°F	0.155	43.197	0.021	486.				
	Gas oil 375 to 650°F	46.810	13,704.213	8.894	2,459.				
	Light gas oil 375 to 500°F	22.140	6,392.557	22.504	3,520.				
	Desulfurized light gas oil	6.405	1,825.942	0.107	58.				
	Hydrocracked jet fuel	12.733	3,616.402	0.003	0.				
	Heavy gas oil	0.716	212.974	0.192	901.				
	Light cycle oil	18.451	5,779.941	10.111	1,749.				
	Blend totals								
90	Jet fuel					22.043	6,282.969	0.613	97.
91	Kerosene					7.226	2,087.194	2.168	1,038.
92	Distillate fuel oil					78.141	23,204.734	39.071	1,683.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

Table J-55. (continued). MASS AND SULFUR BALANCE

Texas Gulf Cluster 1985, Scenario F

Stream number	Process/stream name	Process intake				Process output			
		Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM	Volume,[a] MB/CD	Hydrocarbon weight, Mlbs/CD	Sulfur weight, Mlbs/CD	Sulfur content, PPM
	Blending of olefins								
93	Composition								
	Mixed olefins	2.980	573.283	0.000	0.				
	Light ends	0.745	135.790	0.000	0.				
	Blend totals					3.725	709.073	0.000	0.
	Blending of residual fuel oil								
94	Composition								
	Bottoms 650°+F	2.860	911.127	3.810	4,183.				
	Bottoms 1050°+F	6.371	2,213.643	56.942	25,716.				
	Heavy cycle oil	5.483	1,811.956	9.826	5,422.				
	Blend totals					14.714	4,980.883	70.578	14,169.
	Blending of Lubes								
95	Composition								
	Desulfurized feed	15.474	4,766.188	5.215	1,094.				
	Blend totals					15.474	4,766.188	5.215	1,094.
	Blending of asphalt								
96	Composition								
	Bottoms 1050° +F	1.314	468.560	21.334	45,530.				
	Blend totals					1.314	468.560	21.334	45,530.
	Blending of coke								
97	Composition								
	Coke	3.754	1,444.362	38.550	26,689.				
	Blend totals					3.754	1,444.362	38.550	26,689.

[a] MB/CD except for hydrogen (MMSCF/CD) and refinery gas (FOE MB/CD).

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Table K-1. WEIGHT CONVERSIONS

Unit	Equivalent value
One short ton	2,000 pounds
One metric ton	2,204.6 pounds
One long ton	2,240.0 pounds
One cubic centimeter lead	1.06 grams lead

Table K-2. VOLUME CONVERSIONS

Unit	Equivalent value
One imperial gallon	1.201 U.S. gallons
One liter	0.264 U.S. gallons
One U.S. barrel	42.000 U.S. gallons
One cubic meter	264.173 U.S. gallons
One cubic foot	7.481 U.S. gallons

Table K-3. GRAVITY, WEIGHT AND VOLUME CONVERSIONS FOR PETROLEUM PRODUCTS

(All measurements at 60 Deg F)

Gravity, degrees API	Specific gravity	Gallons per pound	Pounds per gallon	Pounds per barrel	Barrels per short ton	Barrels per metric ton	Barrels per long ton
0	1.0760	0.1116	8.962	376.40	5.31	5.86	5.95
10	1.0000	0.1201	8.328	349.78	5.72	6.30	6.40
15	0.9659	0.1243	8.044	337.85	5.92	6.52	6.63
18	0.9465	0.1269	7.882	331.04	6.04	6.66	6.77
20	0.9340	0.1286	7.778	326.68	6.12	6.75	6.86
22	0.9218	0.1303	7.676	322.39	6.20	6.84	6.95
24	0.9100	0.1320	7.578	318.28	6.28	6.93	7.04
26	0.8984	0.1337	7.481	314.20	6.36	7.02	7.13
28	0.8871	0.1354	7.387	310.25	6.45	7.10	7.22
30	0.8762	0.1371	7.296	306.43	6.53	7.19	7.31
32	0.8654	0.1388	7.206	302.65	6.61	7.28	7.40
34	0.8550	0.1405	7.119	299.00	6.69	7.37	7.49
36	0.8448	0.1422	7.034	295.43	6.77	7.46	7.58
38	0.8348	0.1439	6.951	291.94	6.85	7.55	7.67
40	0.8251	0.1456	6.870	288.54	6.93	7.64	7.76
42	0.8155	0.1473	6.790	285.18	7.01	7.73	7.85
44	0.8063	0.1490	6.713	281.95	7.09	7.82	7.94
46	0.7972	0.1507	6.637	278.75	7.17	7.91	8.04
48	0.7883	0.1524	6.563	275.65	7.26	8.00	8.13
50	0.7796	0.1541	6.490	272.58	7.34	8.09	8.22
55	0.7587	0.1583	6.316	265.27	7.54	8.31	8.44
60	0.7389	0.1626	6.151	258.34	7.74	8.53	8.67
65	0.7201	0.1668	5.994	251.75	7.94	8.76	8.90
70	0.7022	0.1711	5.845	245.49	8.15	8.98	9.12
75	0.6852	0.1753	5.703	239.53	8.35	9.20	9.35
80	0.6690	0.1796	5.568	233.86	8.55	9.43	9.58
85	0.6536	0.1838	5.440	228.48	8.75	9.65	9.80
90	0.6388	0.1881	5.316	223.27	8.96	9.87	10.03
95	0.6247	0.1924	5.199	218.36	9.16	10.10	10.26
100	0.6112	0.1966	5.086	213.61	9.36	10.32	10.49

Table K-4. REPRESENTATIVE WEIGHTS[a] OF PETROLEUM PRODUCTS

Product	Gallons per pound	Pounds per gallon	Pounds per barrel	Barrels per short ton	Barrels per metric ton	Barrels per long ton
Asphalt	0.116	8.60	361	5.54	6.11	6.21
Coke	0.105	9.54	401	4.99	5.50	5.59
Crude petroleum (domestic)	0.142	7.03	295	6.77	7.46	7.58
Crude petroleum (foreign)	0.133	7.50	315	6.35	7.00	7.11
Distillate fuel oil	0.138	7.24	304	6.58	7.25	7.37
Gasoline and naphtha	0.162	6.17	259	7.72	8.51	8.65
Kerosine	0.148	6.75	284	7.05	7.78	7.90
Liquefied petroleum gas	0.221	4.52	190	10.53	11.60	11.79
Lubricating oil	0.133	7.50	315	6.35	7.00	7.11
Residual fuel oil	0.127	7.88	331	6.04	6.66	6.77
Wax	0.150	6.68	280	7.13	7.86	7.99

[a] Approximate or representative figures to be used only for rough estimating. When API or specific gravity is known, Table K-3 should be used.

Table K-5. HEATING VALUES OF CRUDE PETROLEUM AND PETROLEUM PRODUCTS

Item	Gross heating value[a]
Crude petroleum	5,599,100
Petroleum products, average	5,517,000
Dry natural gas	1,021 Btu/cu. ft.
Still gas	6,000,000
Fuel oil equivalent barrel	6,250,000
Natural gasoline	4,620,000
Liquefied gases	4,011,000
Gasoline	5,248,000
Special naphtha	5,248,000
Jet fuel, naphtha-type	5,355,000
Jet fuel, kerosine-type	5,670,000
Kerosine	5,670,000
Distillate fuel oil	5,825,000
Residual fuel oil	6,287,000
Lubricants	6,065,000
Waxes	5,537,000
Petroleum coke	6,024,000
Asphalt	6,636,000

[a] All units in Btu/bbl except as noted

Table K-6. NOMENCLATURE

B/SD Bbls/SD	Barrels per stream day
BTU	British Thermal Unit
cc/gal cc/USG	Cubic centimeters per U.S. gallon
FOE	Fuel oil equivalent
g/gal gm/gal	Grams per gallon
LV	Liquid volume
MB Mbbls	Thousands of barrels
MB/CD	Thousands of barrels per calendar day
MB/SD	Thousands of barrels per stream day
MBPY	Thousands of barrels per year
MKWH	Thousands of kilowatt hours
MKWH/CD	Thousands of kilowatt hours per calendar day
Mlbs	Thousands of pounds
MMB	Millions of barrels
MMB/CD	Millions of barrels per calendar day
MMBPY	Millions of barrels per year
MSCF	Thousands of standard cubic feet
MMSCF	Millions of standard cubic feet
PPM	Parts per million
SCF	Standard cubic feet
$/B/SD	Dollars per barrel per stream day
$MM	Millions of dollars

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APPENDIX L

ALTERNATE FOR REFINERY SO_x CONTROL STUDY

FLUE GAS DESULFURIZATION TECHNOLOGY

A. BACKGROUND

1. COMMERCIAL AND NEAR COMMERCIAL TECHNOLOGIES

With the passage of the Clean Air Act of 1970, considerable effort has been expended by the government and private industry to develop reliable and economical methods for removing SO_2 from the flue gases leaving stationary combustion sources. Many techniques for chemically or physically capturing the SO_2 have been studied and considerable effort is being expended to develop commercially demonstrated processes. The number of SO_2 removal processes currently under development is indeed large. However, it is the intent of this study to identify processes which will be commercially proven and available by 1980. Ideally, this would mean the process would have been installed and operated on at least 100 megawatts prior to this date. A list of the significant SO_2 control processes which have a reasonable chance of meeting these criteria is presented in Table L-1. Included in the table is the current development status of each process and an estimate of the earliest date for commercial availability. In

TABLE L-1

DEVELOPMENT STATUS OF SIGNIFICANT SO_2 CONTROL PROCESSES

As of Mid-1974

PROCESS	REACTANT OR SORBENT	PRODUCT	CURRENT STATUS	Earliest Commercial Availability
Waste Salts				
Limestone Injection and Scrubbing	Limestone	$CaSO_4$	430-Mw plant in operation	1974
Direct Limestone Scrubbing	Limestone	$CaSO_4$	Several plants in operation	1974
Direct Lime Scrubbing	Lime	$CaSO_4$	156-Mw plant operating in Japan	1974
Sodium Solution Scrubbing	Soda ash, caustic soda, trona	Na_2SO_4	250-Mw plant under construction	1974
Double Alkali	Lime (or limestone), soda ash	$CaSO_4$	25-Mw plant planned	1975
Chiyoda	Weak Acid	Gypsum	350-Mw plant under contract	1974
Concentrated SO_2				
Chemico Mag-Ox	Magnesium oxide slurry	10-15% SO_2	150-Mw plant in operation	1975
Stone & Webster/Ionics	Sodium hydroxide solution	100% SO_2	75-Mw plantplanned	1978
Wellman-Powergas	Sodium sulfite solution	100% SO_2	115-Mw plant under construction	1974
Esso/B&W	Unknown dry adsorbent	SO_2/H_2SO_4	Ready for demonstration unit	1979
UOP/Shell	Copper oxide/alumina	SO_2	40-Mw commercial plant operating	1974
TVA	Ammonia	SO_2	Pilot plant operating	1979
FW/Bergbau-Forschung	Dry Char	40% SO_2/H_2SO_4	35-Mw operating	1976
Direct Acid Processes				
Monsanto Cat-Ox	Catalytic oxidation	80% H_2SO_4	110-Mw plant operating	1975
Bergbau-Forschung	Wet Char	20% H_2SO_4	Bench-scale	1976
Elemental Sulfur				
UOP Sulfoxel	Unknown	Sulfur	Pilot-scale operating problems	1979
Atomics International	Molten alkali carbonates	H_2S/Sulfur	Pilot plant operating	1979
Westvaco Char	Char catalyst	Sulfur	Pilot-scale operating	1979
Consolidation Coal	Potassium formate	H_2S/Sulfur	Pilot-scale operating	1979
Bureau of Mines Citrate	Citrate Solution	Sulfur	Ready for demonstration unit	1979
Stauffer Chemical	Sodium phosphate	Sulfur	Pilot operation completed	1979

some cases, the process is presently considered commercialized on oil fired sources. From this list of processes, five candidates were selected for detailed evaluation.

2. INITIAL PROCESS SELECTION

The five processes selected for detailed technical and economic evaluation include:

1. Chiyoda Thoroughbred 101
2. Double Alkali Scrubbing
3. Magnesia Scrubbing
4. Shell Flue Gas Desulfurization
5. Wellman Lord/Davy Powergas Soda Scrubbing

Basic considerations in this selection were the method of sulfur removal, degree of commercialization and form of by-product sulfur.

An attempt was made to cover a variety of SO_2 removal methods while still retaining those processes with a high degree of commercialization. For example, the Chiyoda process absorbent is weak sulfuric acid. Double alkali and Wellman-Lord utilize the sulfite/bisulfite system in different concentration modes. MagOx is a regenerably slurry scrubbing process and, finally, the Shell process represents a dry adsorption process.

In regard to degree of commercialization, all of the above processes have either commercial plants in operation or U.S. demonstration plants planned. Of the five, Chiyoda and Davy Powergas have the most commercial systems operating.

Sulfur product forms represented include throw-away waste salts, concentrated acid and elemental sulfur.

B. DETAILED EVALUATION OF SELECTION PROCESSES

1. BASIS

a. Technical Assumptions

The scope of this study involves the comparison of in process feed and flue gas desulfurization for controlling sulfur emissions from "typical" petroleum refineries. In order to accomplish this objective in a generalized way, certain assumptions were required to facilitate the development of flue gas desulfurization economics. These assumptions and the identification of major sulfur emission sources are presented below.

The major sources of refinery SO_x emissions are shown in Table L-2 along with typical gas conditions for these sources. A detailed breakdown of these major sources together with approximate volumetric flow rates is presented in Table L-3 for the typical refinery locations being studied. These flow rates were determined from the fuel requirements for the various refinery units assuming a gallon of fuel oil generates 2000 scf of flue gas.

TABLE L-2

MAJOR SOURCES OF SOx EMISSIONS IN REFINERIES

1. Sulfur Plant Tail Gas
2. Fluidized Catalytic Cracker Catalyst Regeneration
3. Boiler Plant Stacks
4. Numerous Process Furnace Stacks

Typical Gas Conditions For SOx Emission Source

Source	Claus Plant Tail Gas	FCC Regeneration Gas Before CO Boiler	FCC Regeneration Gas After CO Boiler	Boiler & Furnace Stacks
SOx Concentration, ppmv	12,000-20,000	140-3300(1)	110-2500(2)	1600-2200
Temperature, °F	350	1000-1200	500-800	450-600
Pressure, PSIA	18-20	16-28	14.7	14.7
Dust Loading, gr/scfd	---	0.1-1.4	0.02-1.0	0.1
Approximate Analysis, Mol %				
H_2	3	---	---	
N_2	57	70	72	78
O_2	---	2	4	3
CO	1	8	---	---
CO_2	10	10	11	11
H_2O	26	10	13	8

(1) SO_3 concentration can be 10-60% of total SOx.

(2) Approximately 30% dilution by additional flue gases.

TABLE L-3

REFINERY SULFUR EMISSION SOURCES

Geographical Location	Gulf Coast	East Coast	West Coast	Mid. Cont.
Refinery intake, MB/D	250	150	100	75
STATIONARY SOURCES	Approximate Volumetric Flows, Mscfm (wet)			
Atmospheric Distillation	146.0	87.7	58.3	43.7
Vacuum Distillation	52.7	35.0	23.3	17.7
Naphtha Desulfurization	17.7	10.3	7.3	6.0
Distillate Desulfurization	35.3	20.3	11.7	--
VGO Desulfurization	29.3	17.7	8.7	--
Residum Desulfurization	8.7	13.0	8.7	--
Catalytic Reforming	204.3	102.0	58.3	43.7
Catalytic Cracking	70.0	46.7	29.3	29.3
Hydrocracking	43.7	17.7	17.7	--
Alkylation	52.7	35.0	29.3	29.3
Coking	35.0	14.0	14.0	10.3
H_2 Reformer	73.0	29.3	29.3	--
Steam Boilers	253.3	166.0	105.7	70.0
FLUID CATALYTIC CRACKER				
Regenerator Gas (1)	96.0	64.0	40.0	40.0
Sulfur Plant				
Tailgas before Incineration	18.2	18.2	9.1	1.2

(1) Before CO boiler; after CO boiler volume is 30% greater than figures shown.

Specific assumptions made for developing the off-line economics are:

- Typical stack gas conditions

Temperature, °F	475
Stack pressure, in. wg.	0
Sulfur, ppm	2200
Dust Loading, gr/scf	0.1 max
Moisture, mol %	9
Molecular weight (wet)	28.5

- Base scrubber system capital investment cost based on 100,000 scfm treatment volume and 90% removal of SO_2.

- Base regeneration system capital investment cost estimates based on 200 moles/hr SO_2 recovered equivalent to 77 tons/day of sulfur.

- 50°F of flue gas reheat required for "wet" processes.

- Existing refinery sulfur control assumed to include amine scrubbing unit for sour gas recovery and two-stage Claus unit with incineration.

- FCC regenerator off-gas assumed to have particulate control system and CO boiler.

- Sulfur content of fuel to boilers and fired heaters assumed at 4.5 wt %.

- Plot area available for required flue gas desulfurization equipment.

- Flue gas ducting uneconomical for distances in excess of 2500 feet.

- SO_3 concentration in FCC regenerator flue gas after CO boiler assumed to be similar to that encountered with fossil fuel sources.

- Regeneration of spent absorbent and sulfur recovery system assumed to be centrally located.

This last assumption was made to realize the economies of scale of one large regeneration system as opposed to each scrubbing location having its own regeneration system. However, there are some offsetting costs associated with this concept since regenerated and spent absorbent must be transferred to and from the central system. This requires extra piping and controls. Consequently, an allowance for interfacing the stationary source scrubbers with the central regeneration system was included in the capital investment estimates.

b. Economic Assumptions

The development of capital investment requirements included the following items:

- Other direct costs including site preparation, buildings and service facilities at 10% of process directs.

- Engineering and contractor's fee at 22% of total direct costs.

- Owner's indirects including interest during construction, startup and modifications at 15% of total direct plus engineering costs.

For the preparation of the off-line economic analysis, various unit costs were assumed. These unit costs are summarized in Table L-4. The unit costs incorporated in the refining model, although not exactly these values, are generally in agreement.

For the off-line comparison, maintenance was assumed at 4% of total installed cost (TIC). Fixed cost, including depreciation, return on investment, federal and local taxes and insurance were estimated at 20% of TIC.

TABLE L-4

UNIT COSTS APPLIED IN OFF-LINE ECONOMICS

	$/UNIT
Power, kwh	0.015
Fuel (Low Sulfur), 10^6 Btu	1.25
Steam (High Sulfur Fuel), M lbs	1.25
Reducing Gas (Hydrogen), Mscf (1)	0.70
Water	
Process, M gal	0.40
Cooling, M gal	0.05
BFW, M gal	0.80
Chemicals	
Lime, Ton CaO	22
Limestone, Ton	19
Magnesium Oxide, Ton	119
Soda Ash, Ton	50
Purge Treatment, M gal (2)	1.00
Waste Disposal, Ton (3)	5
Labor (FCOP), yr (4)	65,000

(1) Less fuel credit.

(2) Davy Powergas Process.

(3) Double alkali.

(4) Full coverage operating position.

2. CHIYODA

a. Process Description

The Chiyoda Thoroughbred 101 process for control of SO_x emissions consists of absorbing the sulfur oxides from flue gases in a dilute (2-3%) solution of sulfuric acid. The rich absorbent is then oxidized to produce sulfuric acid from the sulfurous acid present. A portion of the sulfuric acid stream is then reacted with limestone to produce a gypsum by-product.

A schematic process flow sheet for the Chiyoda process is shown in Figure L-1. The flue gas enters a prescrubber where it is cooled to its adiabatic saturation temperature by contacting with the recirculated water. Most of the dust entering the prescrubber is removed upon contact with the liquid which is filtered to remove the solid particles before being returned to the prescrubber. The flue gas then passes to a rectangular, packed absorber where the sulfur oxides are removed from the gas by a dilute sulfuric acid liquor stream.

Much of the sulfurous acid formed by the absorption of SO_2 in the aqueous acid stream is oxidized to sulfuric acid in the presence of the ferric oxide

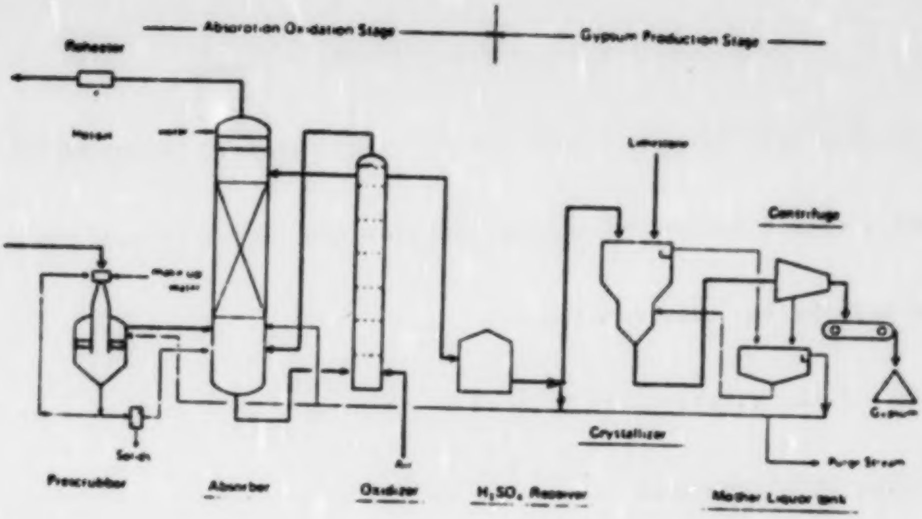
Absorption Oxidation Stage
Gypsum Production Stage
Reheater
Make up water
Limestone
Centrifuge
Gypsum
Crystallizer
Purge Stream
Air
Prescrubber
Absorber
Oxidizer
H_2SO_4 Receiver
Mother Liquor tank

FIGURE L-1

PROCESS FLOW DIAGRAM
CHIYODA THOROUGHBRED 101

catalyst and the residual oxygen in the flue gas. The chemical reactions taking place in the liquid phase in the absorber are:

$$SO_2 + H_2O \rightarrow H_2SO_3$$

$$2FeSO_4 + H_2SO_3 + O_2 \rightarrow Fe_2(SO_4)_3 + H_2O$$

$$Fe_2(SO_4)_3 + H_2SO_3 + H_2O \rightarrow 2FeSO_4 + 2H_2SO_4$$

The absorber effluent liquor passes to an oxidizer column which is a tray tower designed to contact makeup air with the liquor to oxidize any remaining sulfurous acid by the same reactions.

Some of the dilute sulfuric acid is then sent to a crystallizer where it is reacted with limestone to produce gypsum by the equation:

$$CaCO_3 + H_2SO_4 \rightarrow CaSO_4 + H_2O + CO_2\uparrow$$

The gypsum slurry is then centrifuged and the wet gypsum product is removed from the system. The centrate, or mother liquor, is then blended with the remainder of the dilute sulfuric acid stream to redissolve any trace insoluble calcium compounds before returning to the absorber system.

A small purge stream of liquor may be required to keep solubles levels in the system to a minimum. The rate of this purge stream is determined by the rate at which these solubles may enter the system via particulate matter in the flue gas stream or corrosion of the system materials.

b. Process Reliability

The Chiyoda Thoroughbred 101 process is commercially proven. There are presently 12 licensing agreements for the process in Japan, including five plants which are commissioned and operating satisfactorily. The others are at various stages of planning and construction. Of the five operating plants, two are on oil-fired boilers, two are on Claus sulfur plant tail gases and the other on a combination of oil-fired boiler and Claus sulfur plant tail gas. At present, all the Chiyoda licensees are in Japan; consequently, the process has not been commercially demonstrated in the United States. A demonstration plant with a capacity equivalent to 23 megawatts is under construction and will be in operation in Florida in the fall of 1974 to prove the applicability of the process to U.S.

coals. The system will be operating on a utility boiler of Gulf Power Company.

c. Application to Refinery SO_x Control

SO_x removal of up to 95% is possible with the Chiyoda Thoroughbred 101 process. Particulate removal to very low levels is also possible with the liquid to gas ratios used in the prescrubber and absorber. High dust loadings in the feed gas, however, have the undesirable effect of increasing the possibility of pluggage in the packed absorber and increasing solubles levels in the circulating liquor and consequently the impurity level in the by-product gypsum. Chiyoda therefore recommends that an electrostatic precipitator be used for dust loadings in excess of one grain per standard cubic foot. Although most refinery gases which must be treated for SO_x removal have relatively low dust loadings which can approach one grain per standard cubic foot and higher. Since, for the purposes of this study, electrostatic precipitators are assumed to be installed downstream of the FCC regenerator regardless of which desulfurization process is used, the Chiyoda process carries no economic penalty for requiring a relatively low dust loading.

The SO_3 content of the fluid catalytic cracker regeneration gases should pose no problem for the Chiyoda process. The absorption of SO_3 rather than SO_2 is actually preferable as the next process step is to oxidize $SO_2^=$ completely to $SO_3^=$.

d. Capital and Operating Requirements

The capital cost for the gas treating portion of the Chiyoda process is shown in Table L-5 for a gas handling capacity of 100,000 scfm (wet). For the development of these costs, it was assumed that the scrubbing system would be retrofitted to an existing stack and that there is sufficient area to locate the scrubber so that no unusual structural problems would be encountered.

The estimated annual operating requirements for the same gas handling equipment are also shown in Table L-5. For the gas scrubbing system there is little or no difference in capital or operating costs for various SO_x removal efficiencies in the range of 75-95%. Therefore, within the accuracy of these estimates, they can be considered appropriate for any removal efficiency in this range.

TABLE L-5

CHIYODA THOROUGHBRED 101 PROCESS
ESTIMATED CAPITAL COST AND OPERATING REQUIREMENTS

Gas Side--100,000 SCFM

SO_2 Removal Efficiency 90%

Capital Investment ($M)	
Scrubber System	750
Fans, Duct Work and Stack	270
Flue Gas Reheater	60
Process Direct Cost	1,080
Other Directs @ 10%	108
Total Direct Cost	1,188
Engineering and Contractor's Fee @ 22%	261
Subtotal	1,449
Owner's Indirects @ 15%	217
Total Investment	1,666

Operating Requirements

	Annual Usage	($M)
Utilities		
Electric Power	7.7×10^6 kwh	116
Fuel (Reheat)	138×10^9 Btu	173
Water	60×10^6 gal	50
Total Utilities		339

Figure L-2 shows how the capital costs developed in Table L-5 for the base case of 100,000 scfm are extended to the range 30,000 to 200,000 scfm.

Table L-6 shows the development of the capital costs and operating requirements for the liquor side of the Chiyoda process for a liquor capacity equivalent to an SO_x removal rate of 200 lb-moles/hour. Figure L-3 shows the extension of the capital cost data in Table L-6 to the range of 75 to 200 lb-moles/hr.

The value of by-product gypsum in the U.S. is probably less than in Japan because gypsum from natural sources is relatively plentiful and generally considered superior for plasterboard and Portland cement. However, since there are potential markets for the synthetic gypsum, it cannot really be considered a waste salt and the full penalty of landfill cost is also not appropriate. Consequently, it was assumed the gypsum would be available to any seeker willing to pay the hauling cost. Therefore, no product credit nor disposal cost was assigned.

The power costs associated with operating a gypsum production system are relatively insignificant when compared with the requirements for the gas side.

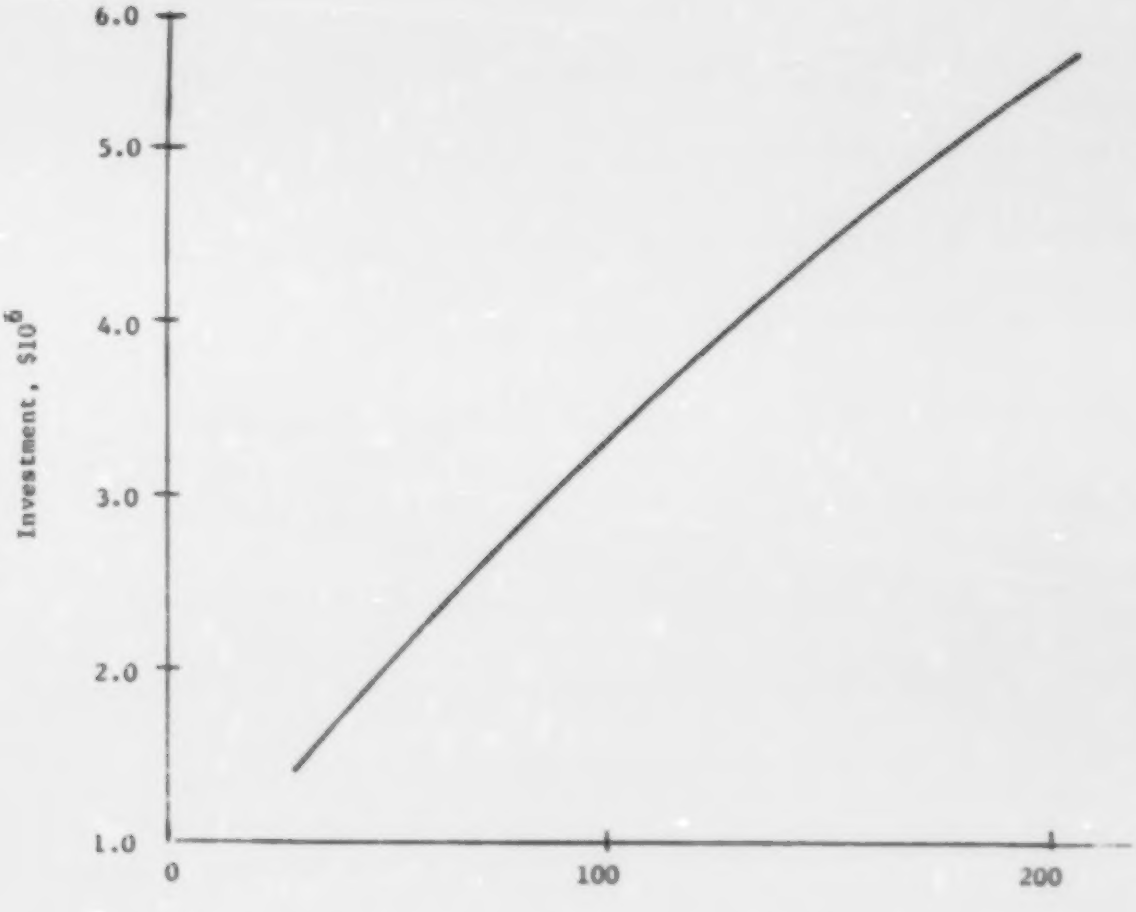

FIGURE L-2

CHIYODA ENGINEERING, CAPITAL INVESTMENT
SCRUBBING SECTION

TABLE L-6

CHIYODA THOROUGHBRED 101 PROCESS

ESTIMATED CAPITAL COST AND OPERATING REQUIREMENTS

Liquor Side--200 lb-mole/hr. SO_2 Removal

Capital Investment ($M)

Material Storage and Feed System	900
Oxidation and Liquor Handling	3,300
Crystallization and Solids Removal	3,400
Interfacing Cost	500
Process Directs	8,100
Other Directs @ 10%	810
Total Directs	8,910
Engineering and Contractor's Fee @ 22%	1,960
Subtotal	10,870
Owner's Indirects @ 15%	1,630
Total Investment	12,500

Operating Requirements

	Annual Usage	($M)
Labor	3 Shift Positions	195
Utilities		
Electric Power $0.015/kwh	20 x 10^6 kwh	
Limestone	86,500T	30
Catalyst	15,400 lb	

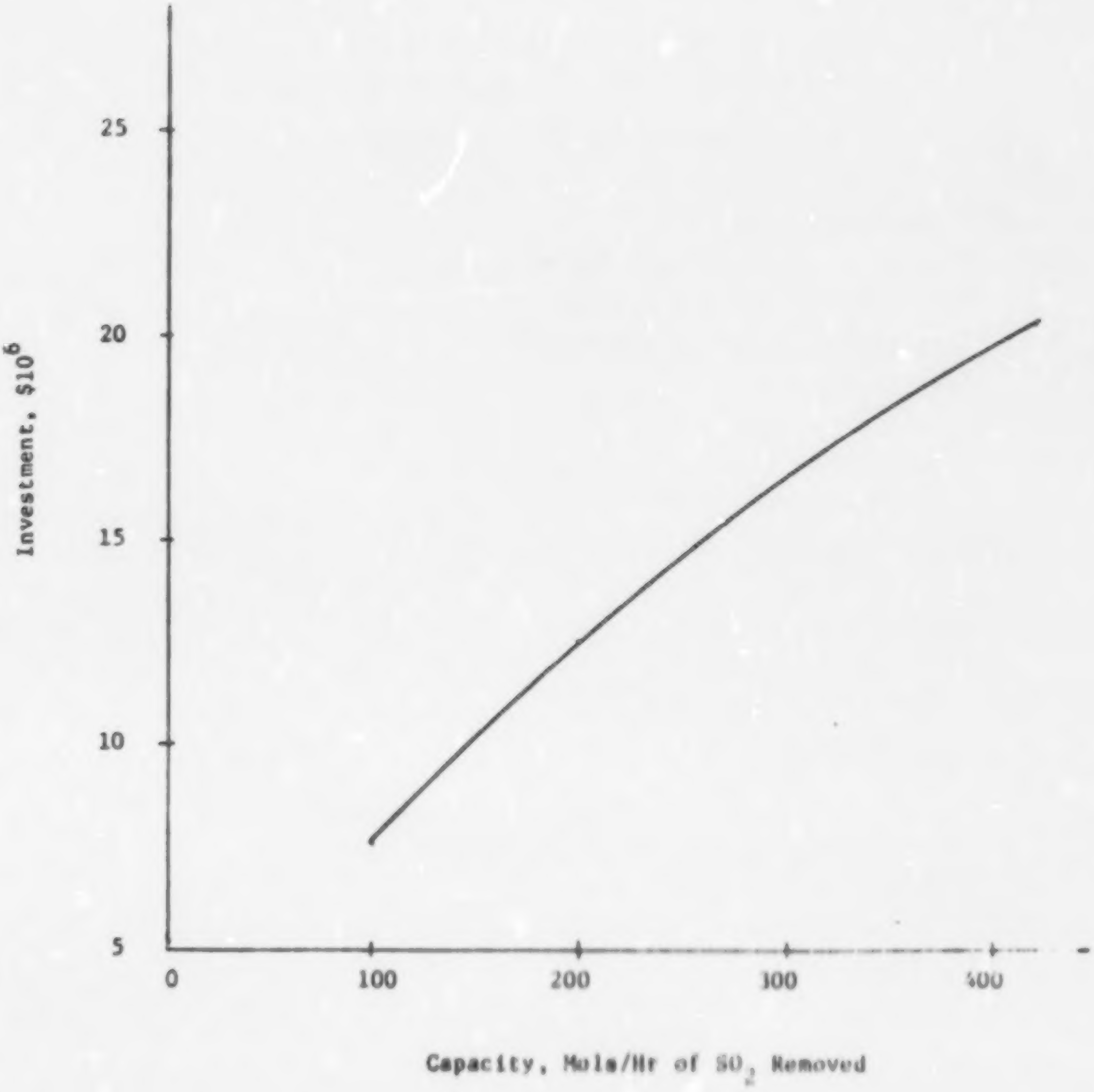

FIGURE L-3

CHIYODA ENGINEERING, CAPITAL INVESTMENT
REGENERATION SECTION

Therefore, no electric power figure is shown in the liquor side operating costs.

The labor costs associated with operating the gas handling system are minimal and are usually considered an extension of the boiler operation. Therefore, operating labor costs are shown only for the operation of the gypsum production system.

3. DUAL ALKALI AND WET LIME SCRUBBING

The economics for dual alkali and lime scrubbing systems are quite similar. Therefore, capital and operating cost estimates were developed for only the dual alkali process and are presented in that section as being approximately typical (total cost basis) for both types of systems. A discussion of lime scrubbing is presented in a separate section.

a. Process Description

In the dual alkali scrubbing process shown schematically in Figure L-4, a mixture of sodium hydroxide and sodium sulfite is used to absorb SO_2 in the

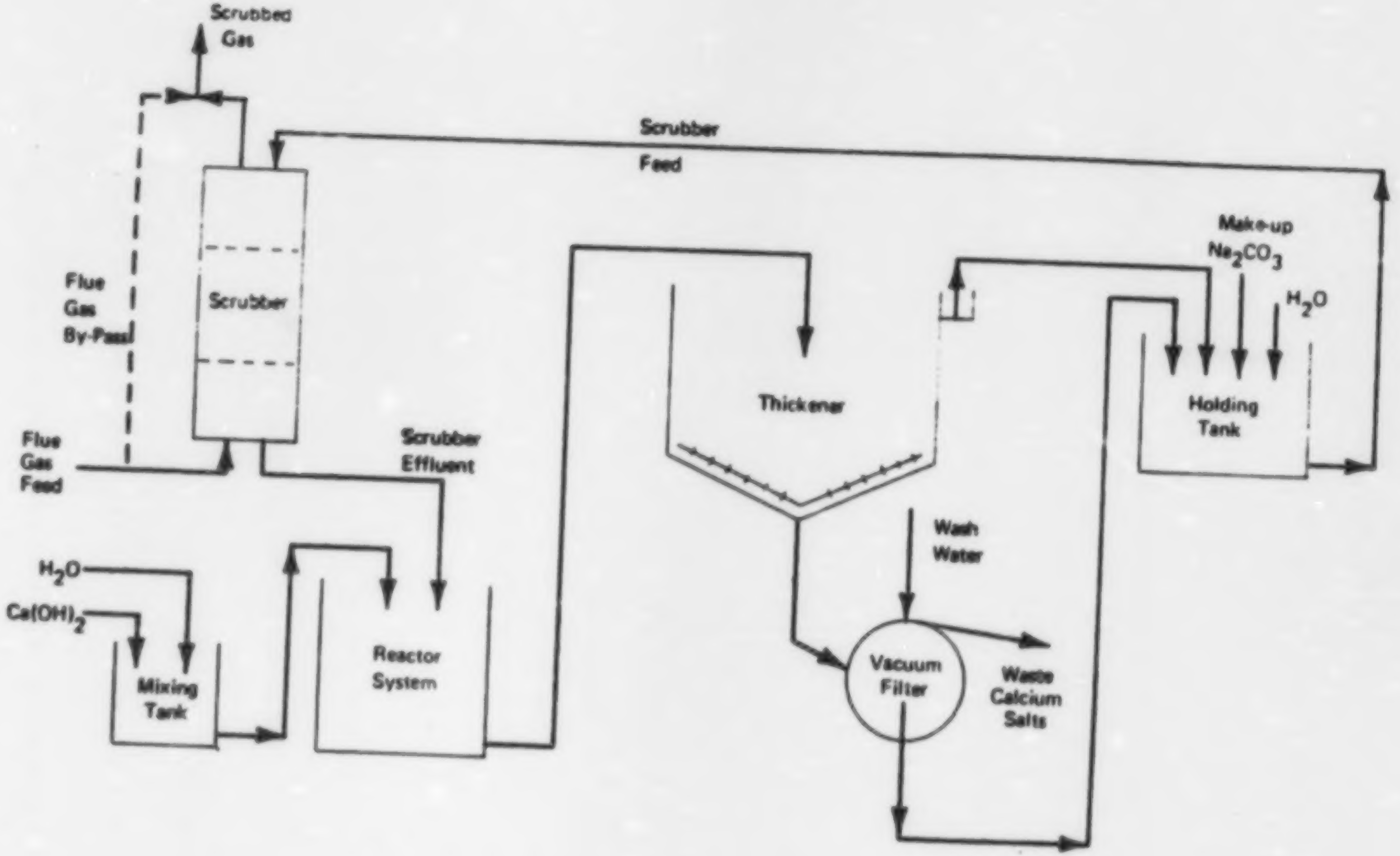
Scrubbed
Gas
Scrubber
Feed
Flue
Gas
By-Pass
Scrubber
Flue
Gas
Feed
Scrubber
Effluent
Make-up
Na_2CO_3
H_2O
Thickener
Holding
Tank
Wash
Water
H_2O
$Ca(OH)_2$
Mixing
Tank
Reactor
System
Vacuum
Filter
Waste
Calcium
Salts

FIGURE L-4 DUAL ALKALI SYSTEM

scrubber system according to the equations:

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3$$

The absorber effluent is reacted with lime in a controlled environment to precipitate calcium sulfite and regenerate the sodium value (as sulfite) which is recycled to the absorber. Some oxidation of sodium sulfite to sodium sulfate does occur in the absorber by the equation:

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4$$

and the lime must also be reacted with sodium sulfate to produce calcium sulfate and active sodium hydroxide. The regeneration equations are:

$$Na_2SO_3 + Ca(OH)_2 \rightarrow CaSO_3 + 2NaOH$$

$$NaHSO_3 + Ca(OH)_2 \rightarrow CaSO_3 + NaOH + H_2O$$

$$Na_2SO_4 + Ca(OH)_2 \rightarrow CaSO_4 + 2NaOH$$

The solid calcium sulfite/sulfate salts are concentrated in the thickener and dewatered on a vacuum filter to produce a waste cake containing 30 to 40% moisture. The process requires a makeup stream of sodium carbonate (or sodium hydroxide) to replace the sodium value lost in the wet solid waste.

b. Process Reliability

The technology of sodium scrubbing to remove SO_x from flue gases has been commercially demonstrated on once-through scrubbing systems. However, regenerating the spent sodium back to an active species for reuse in the scrubber system introduces two problems which are not encountered in the once-through sodium scrubbing scheme. These are: preventing the return of both soluble and insoluble calcium salts to the scrubber system in order to minimize the potential scaling; and regenerating sodium sulfate (which has been formed in the scrubber circuit by absorption of SO_3 or by oxidation of the sodium sulfite). If the sodium sulfate is not regenerated then it must be purged from the system resulting in a liquid waste stream and higher sodium makeup requirements.

These problems have largely been overcome on the pilot plant scale by various researchers using different modes of operation. However, the process has not as yet been commercially demonstrated in the United States. Two dual alkali systems are scheduled for startup in 1974.

c. Application to Refinery SO_x Control

The dual alkali scrubbing process is quite suitable for operation on refinery flue gases. SO_2 removal in excess of 95% can be achieved with capital and operating costs not much higher than for significantly lower removal efficiencies.

The relatively low dust loadings in refinery flue gases can be readily handled by the dual alkali system. Even the fluid catalytic cracker regeneration gases, which have a dust loading of about one grain per standard cubic foot, are relatively lightly loaded when compared with coal-fired boilers for which many dual alkali systems have been proposed. Particulate matter, about 99% of which can be removed from the flue gas in the scrubber system, remains in the liquor as a slurry and is ultimately removed from the system by the vacuum filter as a component in the waste sludge.

The presence of SO_3 in the fluid catalytic cracker regeneration gases increases the concentration of $SO_3^=$ in the circulating scrubbing solution upon absorption, thus aggravating the problem of sulfate regeneration described in the next section.

d. Capital and Operating Requirements

The capital cost for the gas treating portion of the dual alkali system is shown in Table L-7 for a gas handling capacity of 100,000 scfm (wet). For the development of these costs, it was assumed that the scrubbing system would be retrofitted to an existing stack and that there is sufficient area to locate the scrubber so that no unusual structural problems would be encountered. The estimated annual operating requirements for the flue gas treatment system are also shown in Table L-7 There is little or no difference in capital or operating costs for various SO_x removal efficiencies in the range 75-95%. Therefore, these estimated operating requirements can be considered appropriate for any removal efficiency in this range.

Figure L-5 shows how the capital costs developed in Table L-7 for the base case of 100,000 scfm are extended to the range 30,000 to 200,000 scfm.

Table L-8 shows the development of the capital costs and operating requirements for the liquor side of the dual alkali system for a liquor capacity equivalent to an SO_x removal rate of 200 lb-moles/hr. Figure L-6 shows the extension of the capital cost data in Table L-8 to the range of 75 to 200 lb-mole

TABLE L-7

DUAL ALKALI PROCESS
ESTIMATED CAPITAL COST AND OPERATING REQUIREMENTS

Gas Side--100,000 SCFM

SO_x Removal Efficiency 90%

Capital Investment ($M)	
Scrubber System	550
Fans, Duct Work, and Stack	270
Flue Gas Reheater	60
Process Direct Cost	880
Other Directs @ 10%	88
Total Direct Cost	968
Engineering and Contractor's Fee @ 22%	213
Subtotal	1,181
Owner's Indirects @ 15%	177
Total Investment	1,358

Operating Requirements

	Unit Cost	Annual Usage	($M)
Utilities			
Electric Power	$.015/kwh	2.8×10^6 kwh	42,000
Fuel (Reheat)	$1.25/$10^6$Btu	50×10^9 Btu	63,000
Water	$ .80/10^3gal	60×10^6 gal	50,000
Total Utilities			155,000

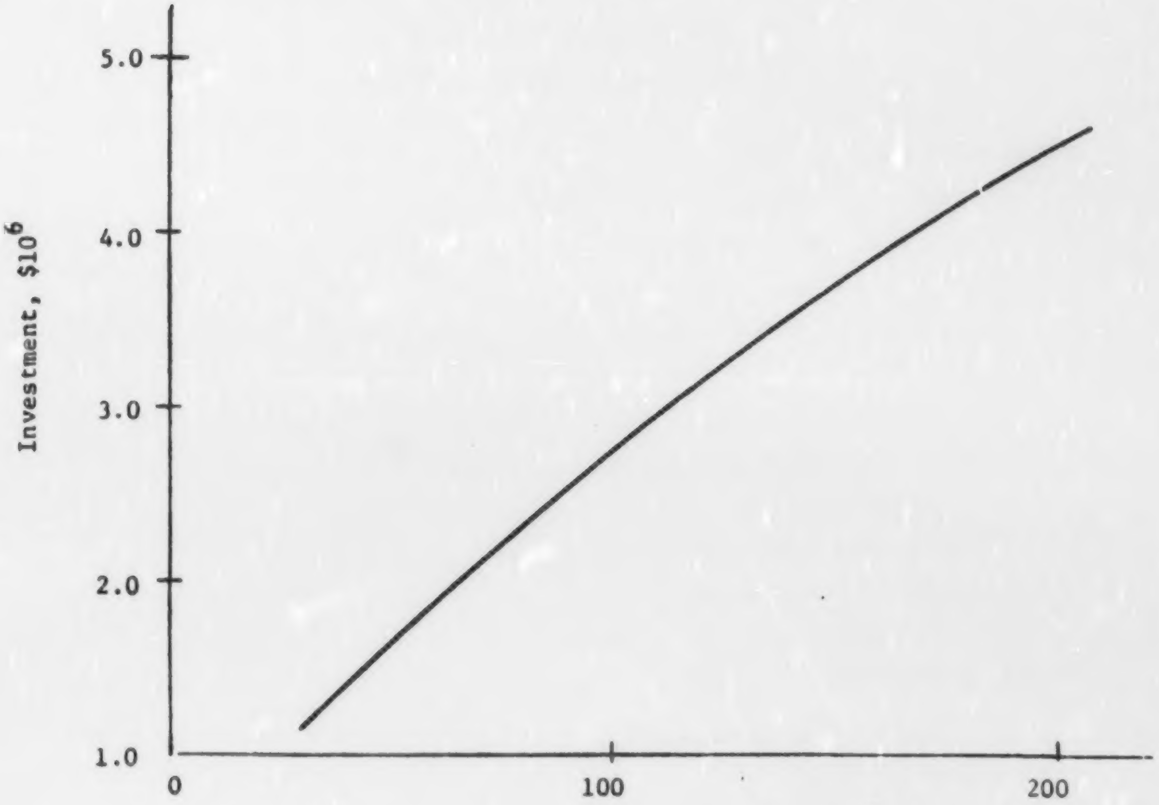

FIGURE L-5

DOUBLE ALKALI, CAPITAL INVESTMENT
SCRUBBING SECTION

TABLE L-8

DUAL ALKALI PROCESS

ESTIMATED CAPITAL AND OPERATING COSTS

Liquor Side--200 lb-mole/hr SO_2 Removal

Capital Cost ($M)	
Material Storage and Feed Systems	1,700
Regeneration System	470
Sludge Thickening and Filtration	770
Interfacing Cost	300
Process Directs	3,240
Other Directs @ 10%	324
Total Directs	3,564
Engineering and Contractor's Fee @ 22%	784
Subtotal	4,348
Owner's Indirects @ 15%	652
Total Investment	5,000

Operating Usages

	Unit Cost	Annual Usage	($M)
Labor	$65,000/Shift Pos.	2 Shift Pos.	130
Utilities			
Elec. Pwr.	$0.015/kwh	1.6×10^6 kwh	24
Soda Ash	$50/Ton	2,650 Tons	100
Lime	$22/Ton	69,700 Tons	1,533
Waste Disposal	$5/Ton	231,000 Tons	1,160

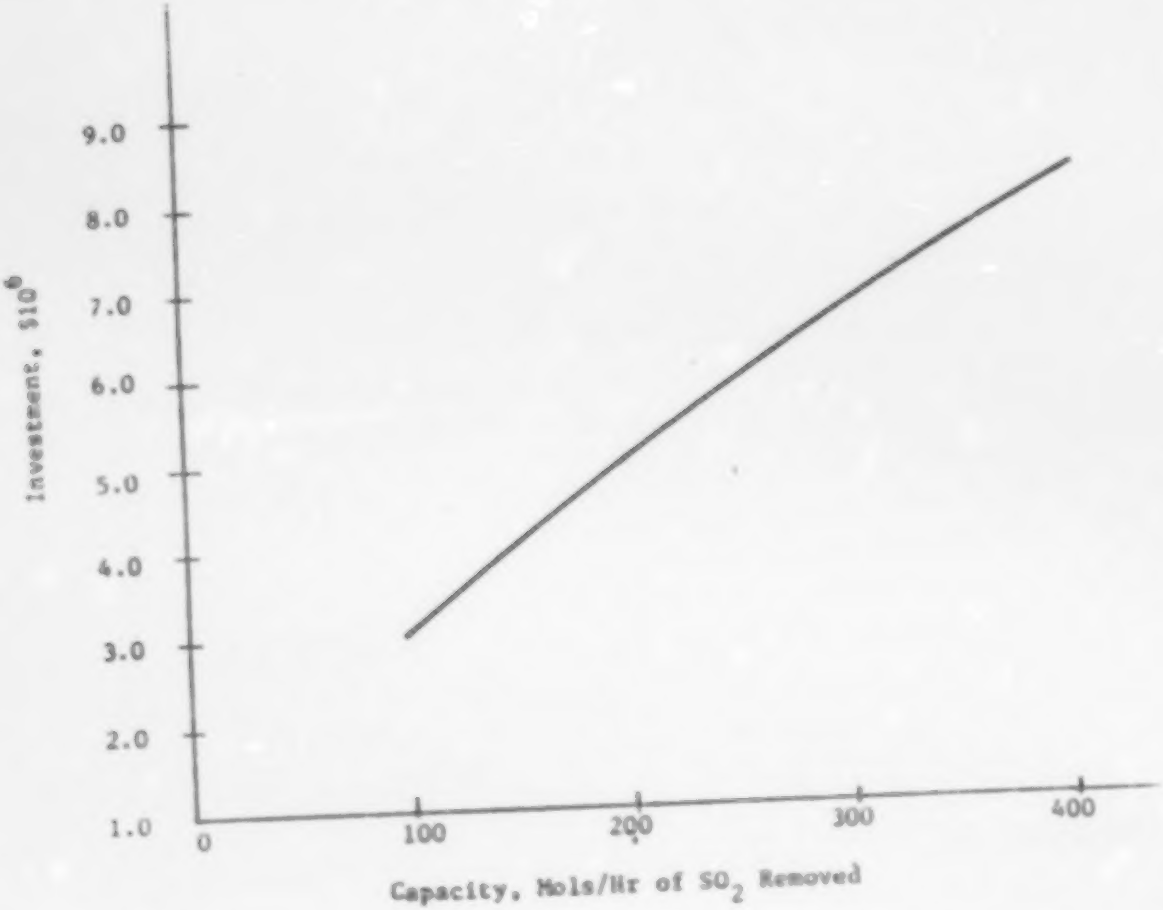

FIGURE L-6
DOUBLE ALKALI, CAPITAL INVESTMENT
REGENERATION SECTION

The power costs associated with operating a regeneration system are relatively insignificant when compared with the requirements for the gas side. Therefore, no electric power figure is shown in the liquor side operating costs.

The labor costs associated with operating the gas handling system are minimal and are usually considered an extension of the boiler operation. Therefore, only operating labor is shown for the operation of the regeneration system. Maintenance costs (at 4% of capital cost) are shown for both the gas and liquor handling systems.

e. Wet Lime Scrubbing

(1) Process Description

In the wet lime scrubbing process, a lime slurry is contacted with the flue gases in scrubbing equipment similar to that used in the dual alkali process to form calcium sulfite by the following reaction:

$$Ca(OH)_2 + SO_2 \rightarrow CaSO_3 + H_2O$$

The scrubber bottom slurry passes to a sludge dewatering system which consists of a thickener and vacuum filter. Lime is added to the clarified water which in turn is recycled to the scrubber system. A schematic flow diagram of this process is shown in Figure L-7.

Either lime or limestone may be used for SO_2 scrubbing. Although the materials cost for limestone is less than for lime, there is a combination of technical advantages to using lime. The reactivity of natural limestone is variable and unpredictable depending on where it is mined. Furthermore, smaller stoichiometric quantities of lime are required for given SO_x removal requirements. For the purposes of this study, the use of lime only is considered.

(2) Process Reliability

The major question regarding the operability of lime scrubbing systems is the potential for scaling and plugging as a result of deposition of solids on the surfaces of the scrubber due either to sticking of slurry solids or crystallization of calcium sulfite or sulfate during the reaction of lime with SO_x.

A second, but less important, problem is the erosion of process equipment and piping which results from the circulation of slurry materials.

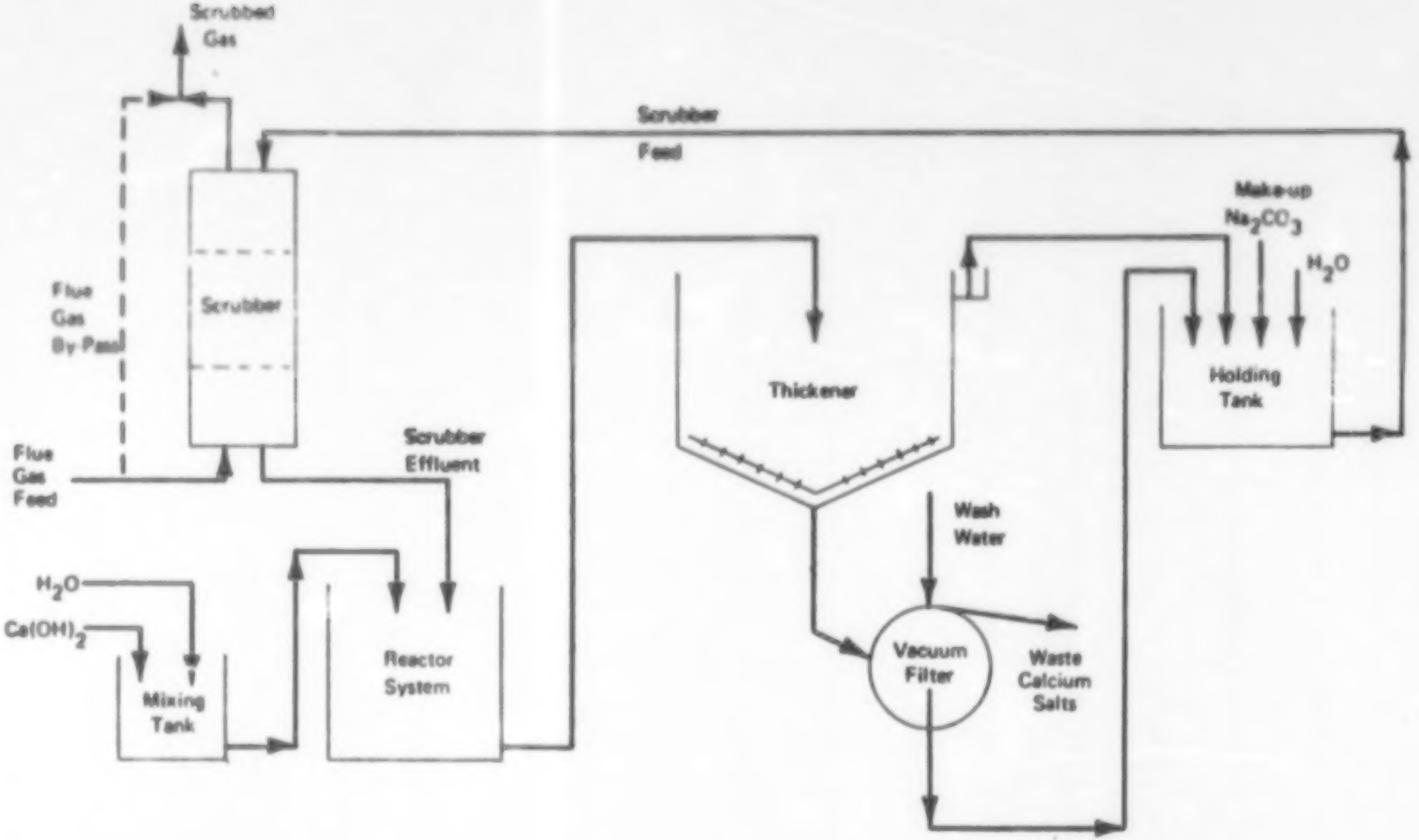
Scrubbed Gas
Scrubber Feed
Flue Gas By-Pass
Scrubber
Flue Gas Feed
Scrubber Effluent
Make-up Na_2CO_3
H_2O
Thickener
Holding Tank
Wash Water
H_2O
$Ca(OH)_2$
Mixing Tank
Reactor System
Vacuum Filter
Waste Calcium Salts

FIGURE L-7 DUAL ALKALI SCRUBBING WITH LIME REGENERATION

Until recently no concensus had been reached regarding the optimal operating mode for lime/limestone systems as the problems of scaling, plugging and erosion always arose. However, since April 1973 the SO_x control system (lime scrubbing) at Louisville Gas and Electric's Paddy's Run Station No. 6 in Kentucky has been operating satisfactorily with SO_x removal at about 90% and no evidence of scaling, plugging or erosion. The plant operated for periods up to twelve days. Of the several lime/limestone systems which have been operated, Paddy's Run appears to offer the most encouragement for the potential long-range operation of a wet lime scrubbing system.

(3) Applicability to Refinery SO_x Control

Despite some claims of up to 98% SO_2 removal capability with lime scrubbing, the ability to maintain such high removal rates for extended period of operation has not been demonstrated and is still questionable.

Since the scrubber system is designed to handle a slurry, dust loads characteristic of refinery flue gases including fluid catalytic cracker regeneration gas would be expected to pose no problem. These solids would be removed from the system along with the by-product waste sludge.

The major question regarding the operability of lime scrubbing systems is the potential for scaling and plugging of the scrubber system. The process has been piloted by many people and there has been no concensus as to the optimum operation.

It is anticipated that in a refinery, several gas treating units (scrubbers) will be located near the emissions sources and the sulfur-rich liquor would be piped to a common liquor handling or treatment facility. Since slurries must be transported between the gas-treatment and liquor handling systems to and from the scrubbers, a serious turndown problem arises. The solids must be maintained in suspension by keeping a minimum velocity in slurry pipelines. However, when several scrubbers are down, the flow through the header will decrease below this minimum. Therefore, an additional flow must be circulated and another pipeline must be provided to carry the extra flow back to the liquor system. This results in extra power consumption for pumping and additional capital investment for installation of long pipelines. Furthermore, the main headers must be larger than would otherwise be necessary because they must carry the recirculating slurry in addition to the primary flow.

SO_3 in the fluid catalytic cracker regeneration gases will result in a waste solid relatively rich in calcium sulfate which is no less environmentally acceptable. Process-wise, higher sulfate levels in the scrubber circuit are of no concern.

4. MAGNESIA SCRUBBING

a. Process Description

The MagOx process, developed by Chemical Construction Company and Basic Chemicals, removes sulfur oxides from flue gases by scrubbing with a recirculating slurry consisting of magnesia crystals and some insoluble particulates. Better than 90% removal efficiency of SO_2 has been demonstrated at Boston Edison using this process.(8) The major by-product of the magnesium oxide scrubbing process is a dilute (12-16%) SO_2 gas stream which is usually converted into sulfuric acid. The overall process is divided into the following operations:

1. SO_2 absorption
2. Slurry processing
   a. Contamination control
   b. Hydrate conversion
3. Dewatering

4. Drying
5. Calcining
6. Slurry makeup
7. Sulfuric acid production

The manufacture of sulfuric acid is well known technology; therefore, this process is not described in detail. It is, however, included in the process flow diagram.

The process chemistry for Steps 1, 4 and 5 above is given below.

Absorption

Main Reactions

$$MgO + SO_2 + 3H_2O \rightarrow MgSO_3 \cdot 3H_2O$$

$$MgO + SO_2 + 6H_2O \rightarrow MgSO_3 \cdot 6H_2O$$

Side Reactions

$$MgSO_3 + SO_2 + H_2O \rightarrow Mg(HSO_3)_2$$

$$Mg(HSO_3)_2 + MgO \rightarrow 2MgSO_3 + H_2O$$

$$MgO + SO_3 + 7H_2O \rightarrow MgSO_4 \cdot 7H_2O$$

Drying

Main Reactions

$$MgSO_3 \cdot 3H_2O \xrightarrow{\Delta} MgSO_3 + 3H_2O$$

$$MgSO_3 \cdot 6H_2O \xrightarrow{\Delta} MgSO_3 + 6H_2O$$

$$MgSO_4 \cdot 7H_2O \xrightarrow{\Delta} MgSO_4 + 7H_2O$$

Side Reaction

$$MgSO_3 + 1/2\ O_2 \xrightarrow{\Delta} MgSO_4$$

Regeneration

$$MgSO_3 \xrightarrow{\Delta} MgO + SO_2$$

$$MgSO_4 + 1/2\ C \xrightarrow{\Delta} MgO + 1/2\ CO_2 + SO_2$$

(1) SO_2 Absorption

The flue gas enters the venturi absorber as shown in Figure 1.-8. There it is contacted counter-currently with the recycled slurry containing magnesium oxide (MgO), magnesium sulfite ($MgSO_3$) and magnesium sulfate ($MgSO_4$). The slurry also contains particulates. As shown in the process chemistry, sulfur dioxide is absorbed into the slurry by magnesium oxide to form magnesium sulfite and some

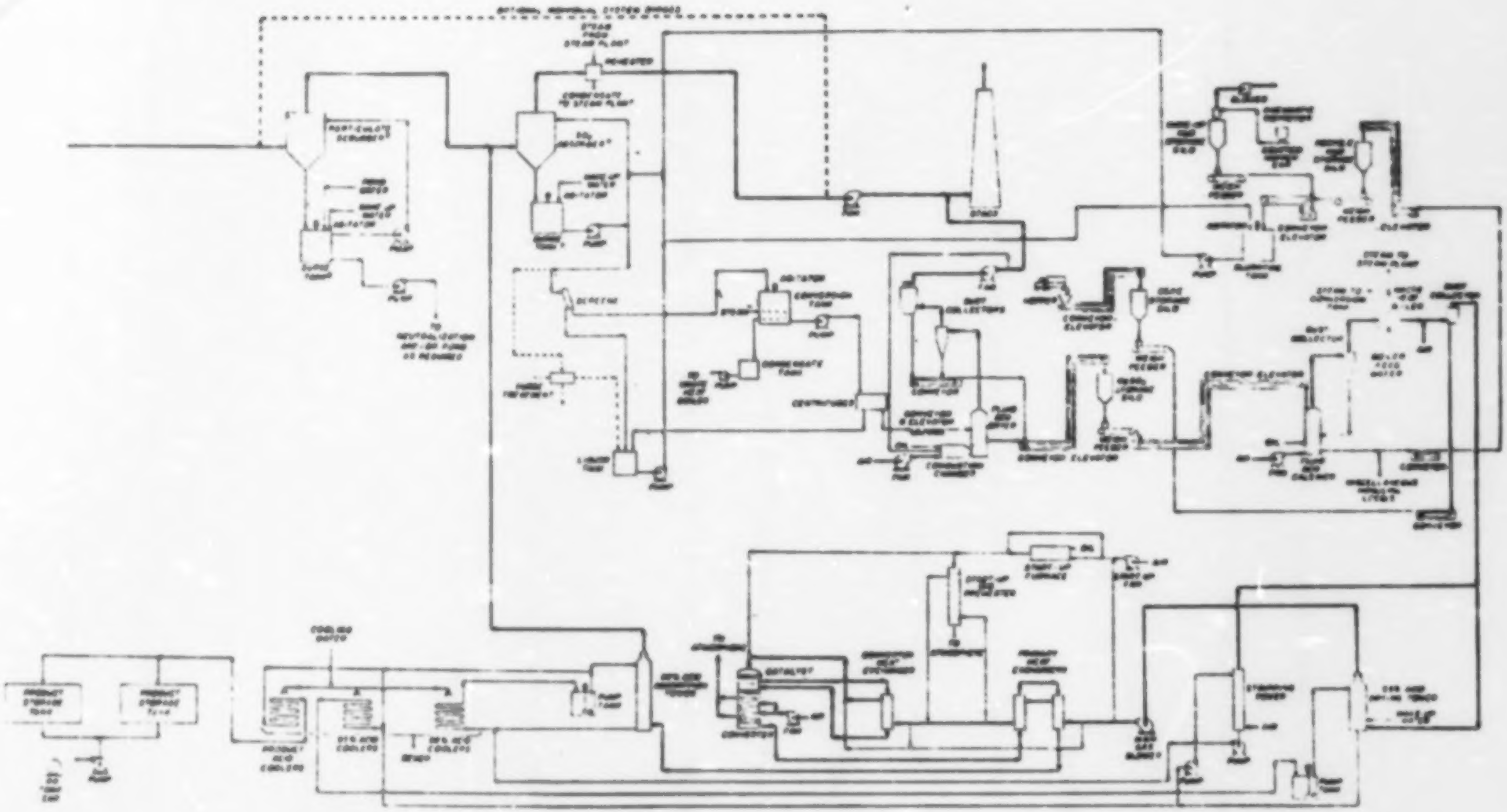

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Figure L-8 Flow Diagram— Magnesia Slurry Scrubbing-Regeneration magnesium bisulfite. To a lesser degree, other reactions occur between the flue gas and the magnesium compounds in the recycled slurry, including reactions producing magnesium sulfate.

(2) Slurry Processing

The slurry processing operation can be subdivided into two stages: contamination control and hydrate conversion.

Contamination Control

When this process is applied to oil-fired sources which generate a low particulate flue gas (0.0228 gr/scfd was measured at the Boston Edison pilot plant based on #6 fuel oil)(8), separate particulate removal is not required. On coal-fired sources, separate particulate scrubber and ash pond or electrostatic precipitators are employed to reduce the particulate loading before the flue gas enters the absorber. Even with oil-fired systems, there will be some buildup of insoluble particulates and soluble contaminant in the system. Some means of purging the system of these solids must be provided. With oil firing, the buildup of insolubles is small since there are unavoidable losses of both particulates and magnesium solids in the process of drying, calcining and conveying solids within the plant.

However, because of the regenerative and cyclic nature of the magnesium oxide scrubbing process, it is necessary to minimize the amount of contaminants which build up in the sulfur dioxide scrubbing loop.

Both soluble and insoluble impurities build up in the sulfur dioxide scrubber loop. Soluble impurities enter the system in the makeup water, principally, and to a lesser degree with makeup MgO. The major soluble contaminants expected are calcium oxide (CaO), sulfate ion ($SO_4^=$) and chloride ion (Cl^-).

The major source of insoluble contaminants is fly ash. For flue gas from oil-fired units, the rate of fly ash addition is low. Even so, the fly ash mass rate of addition to the scrubber loop is greater than the addition rate of soluble contaminants by a factor of about 200 for oil-fired sources.

One method of contamination control is to react a purge stream from the recirculating slurry with sulfur dioxide in a solubilizing tank. Here the magnesium compounds of sulfite and sulfate would be solubilized followed by filtration to remove fly ash as a cake. The magnesium compounds in the filtrate can be precipitated by adding MgO to the filtrate in a separate reaction vessel. By filtration, the magnesium solids can be recovered and returned to the system, with the contaminated filtrate discarded to an evaporation pond.

Hydrate Conversion

Reported pilot plant data for MgO slurry scrubbing indicate that $MgSO_3 \cdot 3H_2O$ can be obtained from the hexahydrate by thermal conversion. In the process flow diagram, the particular sequence of size screening, thermal conversion, dewatering and drying was chosen because:

1. Crystals of the hexahydrate require more heat for drying than crystals of the trihydrate.
2. Trihydrate crystals are smaller than hexahydrate crystals and are more difficult to dewater.
3. Net heat savings can result from thermally converting crystals in a thickened slurry before drying. Thickened slurries have the fastest conversion rates and require less sensible heat.

Returning to Figure L-8, a bleed stream of the absorbent slurry is withdrawn for regeneration of magnesium sulfite and magnesium sulfate to magnesium oxide. To separate the larger magnesium sulfate hexahydrate crystals in a thickened slurry, wet screens are utilized. The return stream to the scrubber loop consists of a dilute slurry containing the smaller crystals. These smaller crystals can serve as nuclei for further crystal growth in the scrubber loop. A thickened slurry of about 40% solids from the wet screens is fed to a thermal conversion tank where heat is added to convert $MgSO_3 \cdot 6H_2O$ to $MgSO_3 \cdot 3H_2O$.

(3) Dewatering

The slurry stream from the thermal conversion tank containing the hydrated magnesium crystals, both as a slurry and as a saturated solution, is discharged to a continuous centrifuge. There, partial dewatering produces a cake containing a mixture of crystals of $MgSO_3 \cdot 3H_2O$, $MgSO_3 \cdot 6H_2O$, $MgSO_4 \cdot 7H_2O$ and unreacted MgO. However, during the thermal conversion treatment, most of the sulfite crystals convert to the trihydrate form. The clear, saturated, centrate liquor is returned to the main recirculating slurry line. The solids cake, containing about 15% water, is conveyed to a fluid bed dryer.

(4) Drying

In a separate combustion chamber, fuel oil is burned with a minimum of excess air and fed to a fluid bed dryer. By direct contact the drying gas vaporizes both the bound moisture (water of hydration) and the remaining free moisture in the solids feed under non-oxidizing conditions to give an essentially anhydrous solids product (less than 3% moisture). About one-third of the total exhaust gas from the dryer is recycled to the combustion chamber to minimize heat losses and to increase the superficial gas velocity in the dryer.

The flue gas from the dryer contains particles of magnesium and fly ash which must be removed before being sent to the stack. For the exhaust temperature of 400°F, a cyclone and bag filter combination can be used to return these crystals to the dryer product conveyor. The cleaned dryer flue gas is combined with the outlet flue gas from the SO_2 absorber.

(5) Calcining

The crystals from the dryer are fed via conveyor to a fluid bed calciner where they are contacted with a hot flue gas to regenerate MgO. The flue gas is generated by direct combustion of fuel oil with 5% excess air. $MgSO_3$ is converted directly to MgO, and $MgSO_4$ is converted to MgO by feeding coke to the calciner along with the dried crystal feed. Typically, calciner feed crystals contain 5-10% $MgSO_4$, the remainder being essentially all $MgSO_3$.

Off-gas from the calciner contains 12-16% SO_2 by volume which is liberated from MgO regeneration. The SO_2-rich flue gas from the calciner is sent to the sulfuric acid plant where 98% H_2SO_4 is made. (Production of 98% concentrated sulfuric acid was chosen for this process.)

The flue gas contains a significant amount of magnesium crystals which must be collected. However, due to the high flue gas exit temperature of 1600°F, a bag filter cannot be used without first cooling the flue gas. Therefore, a hot cyclone is used primarily on the flue gas followed by a heat recovery unit which lowers the flue gas temperature to 700°F. If instead the flue gas temperature is lowered to the proper operating range for a bag filter of 425-500°F at the heat recovery unit, there would be a chance of sulfuric acid mist condensation in the bag filter. Although this is only a remote chance, air is added between the waste heat boiler and the bag filter to reduce the flue gas temperature from 700°F to a temperature suitable for bag filter operation.

(6) Slurry Makeup

The regenerated MgO crystals from the calciner are conveyed to a makeup tank where they are re-slurried along with fresh MgO makeup crystals which are added to replace miscellaneous magnesium losses throughout the system. A bleed stream of recycle slurry from the venturi scrubber loop is used as liquor in the makeup tank and the new slurry is returned to the absorption system.

(7) Sulfuric Acid Production

Most of the more recent sulfuric acid plant designs utilize the contact process in which the following sequence takes place:

a. The generation of a sulfur-dioxide-rich gas from an appropriate raw material.
b. Cooling, purification and drying of the gas.
c. Reheating of the gas to the proper temperature for conversion to sulfur trioxide (SO_3).
d. Catalytic oxidation of SO_2 to SO_3.
e. Cooling of the SO_3-containing gas.
f. Production of sulfuric acid by absorption of SO_3 in concentrated sulfuric acid.

The hot flue gas from the calciner is fed to the sulfuric acid plant after cleansing by the bag filter. Utilizing a bag filter to clean the flue gas instead of a wet scrubber is referred to as a "dry" system in the sulfuric acid production process. One advantage of the "dry" system is that it cleans the calciner off-gas without lowering its temperature. This reduces the amount of reheat necessary to increase the flue gas temperature to 830°F as required by the converter in the sulfuric acid process. A second advantage is that the moisture content of the feed gas must be below the mole ratio of 1.11, $H_2O:SO_2$ required to make 98% H_2SO_4. If the water content of the gas is below 4.4 wt. % (which corresponds to the proper ratio of H_2O to SO_2 for this flue gas), additional water can be added at the sulfuric acid plant to insure the proper ratio. In a "dry" system this is possible; in a "wet" system a more difficult water removal step would have to be implemented in the sulfuric acid plant.

Air must be bled into an SO_2-containing flue gas stream before being fed to a sulfuric acid unit to obtain an $O_2:SO_2$ mole ratio of 1.4. This addition lowers the SO_2 percent in the gas stream but it is necessary to obtain efficient conversion of SO_2 to SO_3. On this basis, the percent SO_2 in the flue gas from the calciner is roughly 10% lower than the percent SO_2 in the off-gas from a conventional sulfur burner used to generate feed gas for the sulfuric acid process.

Another problem that sulfuric acid plants must be concerned with is their gaseous sulfur pollutants emitted into the atmosphere. EPA emission standards for sulfuric acid plants allow a maximum of 4.0 lbs-SO_2/ton and 0.15 H_2SO_4 mist per ton H_2SO_4 emissions. This translates to a minimum of 99.7% required conversion of inlet SO_2 to H_2SO_4 in the sulfuric acid plant. By recycling tail gas from the sulfuric acid plant to the stack gas entering the venturi absorber, an overall efficiency of the sulfuric acid process greater than 99.7% can be obtained.

b. Process Reliability

Good operating results from actual tests that have been performed on the magnesium oxide scrubbing process are minimal. To date, no commercial units have been built utilizing this process, although large-scale demonstration units at both utility plant and industrial sites have been operated. Our concensus, based on available data, is that:

- The process is feasible.
- There are many variations of the process, giving it the flexibility required to meet a wide variety of commercial needs.

- All of the "bugs" have not been worked out for any of these processes.

- The processes that have been attempted on a demonstration scale have not incorporated the best possible technology for some of the operations characterizing the overall process.

- The economics of this process depend on the ability to utilize a dilute (12-16%) byproduct gas stream of sulfur dioxide.

- More demonstration tests are being conducted to develop basic design data and operating experience.

Development work on the magnesium oxide slurry scrubbing process has been performed by Chemico-Basic companies and the Babcock & Wilcox Company in the United States, Russia and Japan. In addition, the Grillo-Werke Company of the United States has tested the magnesium oxide-manganese dioxide slurry scrubbing variation of the process.

Although data are not available for filtration of hexahydrate crystals, vendors indicate that a moisture content of 10-20% could be expected compared with about 5% moisture after centrifugation.

In the venturi scrubbing and recycle loop section of this process, suitable demonstrations have been made at a corresponding high degree of reliability. Design provisions have to be made to prevent problems of corrosion and erosion associated with low pH slurry scrubbing. Also, spare pumps can be provided to circumvent shutdowns and minimize problems due to possible slurry settling in the lines. Proper wash facilities can be provided for slurry lines.

The slurry treatment process mentioned previously has not been demonstrated in its entirety, although experts generally agree that this arrangement is feasible. Chemico-Basic has shown satisfactory performance in centrifuging a magnesium sulfite hexahydrate slurry at their 150 megawatt demonstration unit at the Boston Edison Company. Similar tests for the trihydrate crystals have not been performed.

Pilot plant results have not been obtained for thermal conversion of magnesium sulfite hexahydrate crystals to the trihydrate form. Kinetics indicate that conversion should go rapidly and completely under proper conditions.

One area of slurry treatment that is lacking in data is treatment of a bleed stream from the absorber loop for contamination control. The nature of this problem and possible solutions have been described earlier in this section. This could be a critical area for some variations of the magnesium scrubbing process utilizing a high fly ash loading or a mineral-rich makeup water stream.

Chemico-Basic has demonstrated rotary dryer performance where the feed material was predominantly a magnesium sulfite hexahydrate cake. The dryer product contained 3% moisture when operated within a temperature range of 600-800°F.(7) Reports indicate considerable difficulties associated with the co-current, rotary dryer operation. One type of problem is associated with the excessive dust entrainment from the system. This problem was primarily due to poor control of gas velocity through the dryer. The more serious problem associated with the rotary dryer involved agglomeration of dryer feed cake and insufficient drying within these masses. Downtime in the dryer was significant, causing the dryer to be the bottleneck in the process. It has been suggested that either a countercurrent dryer or a fluid bed dryer be utilized. The key to making this process work would be to produce a uniform centrifuge cake with low water content; to introduce the feed in a discrete, steady manner; to maintain good control of the superficial gas velocity through the fluid bed; and to incorporate a highly efficient dust removal system.

As in the drying operation, Chemico has also demonstrated rotary calciner operation. Nearly complete conversion was obtained between 1600-2000°F with the addition of coke.(7) It should be mentioned, however, that problems were initially encountered in calcining the product magnesium oxide at excessive temperatures. The product was burned and rendered inactive. Satisfactory performance has been demonstrated in Japan with fluid bed calciners. These results were obtained at 1800°F without the addition of reducing coke. The process (e.g., lower temperatures) should proceed even better with the addition of coke.

C. Application to Refinery SO_x Control

There are few special provisions associated with the magnesium oxide slurry scrubbing process that would be necessary for its adaptation to a refinery. The amenability of this process to a refinery can be considered under three categories:

a) Space Requirement of Process
b) Degree of Centralization of Process
c) Process Dependence on Flue Gas Composition

At a refinery site there may be many sources of sizeable flue gas emissions in need of sulfur controls. If it is impractical, infeasible or impossible to duct two or more of these outlet flues to a common scrubber unit, separate scrubbers would have to be installed at each flue gas source.

The amount of space required for the regeneration section of the magnesium process would depend on the degree to which this process is centralized. If each scrubber within the refinery has its own slurry treatment and calcination unit, this case would represent the maximum space requirement. If one centralized slurry treatment system, drying and calcination unit is utilized, this would represent the best case with the minimum space requirement for this process.

The applicability of this process to a refinery would depend on the refinery's ability to utilize the dilute sulfur dioxide byproduct gas stream. The refinery would either have to be capable of selling the sulfur dioxide in the gas stream to a neighboring plant producing a sulfur product, or it would have to build its own sulfuric acid, elemental sulfur, or sulfur dioxide manufacturing plant. The remaining alternative would be for the refinery to truck the dried magnesium sulfite/sulfate crystals to a nearby sulfur-utilizing plant. A calcination unit for the liberation of the sulfur dioxide in a flue gas stream could be built at this plant. Regenerated magnesium oxide crystals could then be trucked back to the refinery for addition to the venturi scrubbers which would conceivably be spread out over the refinery.

The applicability of the magnesium scrubbing process to a refinery will also depend on the level of sulfur dioxide and particulates in the inlet flue gas. There are no data to indicate that varied sulfur dioxide levels in the inlet flue would be difficult to remove by magnesium oxide slurry scrubbing. If, however, a high particulate loading exists in the inlet flue, as it would in the flue gas from a coal-fired boiler or from a fluidized catalytic cracker unit (FCC), then a separate particulate removal device would be necessary upstream of the sulfur dioxide venturi absorber. Either an electrostatic precipitator or a separate particulate scrubber (preferable) could be installed to remove the particulates. If a wet scrubber is used, an ash pond would be necessary for the particulate that is removed from the flue gas. If this were not feasible, it would be necessary to provide for disposal of the fly ash.

The potentially high concentrations of SO_3 in FCC catalyst regenerator gas would result in higher magnesium sulfate concentrations in the circulating slurry. This is less of a problem with the magnesium system, since magnesium sulfate can be regenerated to magnesium oxide as previously explained.

d. Capital and Operating Requirements

Estimated capital investment and operating requirements for the Mag-Ox process are given in Tables L-9 and L-10. The first table presents the costs associated with the scrubber section of the process and the second table contains costs for the combined regeneration section and sulfuric acid plant. Investment as a function of capacity for scrubbing and regeneration is presented in Figures L-9 and L-10.

The operating requirements in Table L-9 do not take account of provisions for slurry treatment for contamination control or for any specific type of particulate removal devices. (Some particulates are removed by the venturi absorber.) In addition, it is assumed that regeneration occurs on-site.

Fuel requirements are based on 50°F of flue gas reheat. This is in addition to the temperature rise due to compression of the flue gas through the induced draft fan downstream of the scrubber, and the temperature rise due to dryer off-gas addition to the scrubber flue gas.

TABLE L-9

CAPITAL & OPERATING REQUIREMENTS

Magnesium Oxide Scrubbing System

Basis
- 100,000 scfm Flue Gas
- 2,200 ppm SO_2
- 90% SO_2 Removal
- 8,000 Operating Hours/Year

Estimated Capital Investment

	Cost ($M)
Scrubber System	435
Fan, Duct Work and Stack	270
Gas Reheat System	60
Process Directs	765
Other Directs @ 10%	77
Total Directs	842
Engineering and Contractor's Fee @ 22%	185
Subtotal	1,027
Owner's Indirects @ 15%	154
Total Investment	1,181

Annual Operating Requirements

	Usage	Unit Cost	Annual Cost ($M)
Labor	4 Man Years	$16,000/man yr	64.0
Utilities:			
Fuel	1029 BTU/M scf	$1.25/MM BTU	63.0
Process Wtr	68×10^6 gal/yr/10^5 scfm	$0.40/M gal	27.2
Power	6.25 kw/M scfm	$0.015/kwh	75.0
Total Utilities			165.2

TABLE L-10

CAPITAL & OPERATING REQUIREMENTS

Magnesium Oxide Regeneration System

Basis: 200 # Moles SO_2 Treated/Hr
(240 Ton/Day of 98% H_2SO_4)

Estimated Capital Investment	Cost ($M)
M_gO Regeneration (Dewatering, Drying & Calcining)	3.230
Sulfuric Acid Plant (Contact Unit & Storage)	1.910
Purge Treatment System	300
Allowance for Interfacing	500
Process Directs	5440
Other Directs	544
Total Directs	5984
Engineering and Contractor's Fee @ 22%	1316
Subtotal	7300
Owner's Indirect @ 15%	1095
Total Investment Cost	8395

Annual Operating Requirements

H_2SO_4 Plant + Regeneration System Variable Costs:	Usage	Unit Cost	Annual Cost($M)
Labor	12 Man Years	$16,000/ManYr	192.00
Utilities:			
Fuel	47.5 gal/T acid(1)	$1.25/MMBtu	647.42
Water-Boiler Feed	181 gal/T acid	$0.80/Mgal	11.30
-Cooling	17.7 Mgal/T acid	$0.05/Mgal	69.32
-Process	4.2 gal/T acid	$0.40/Mgal	0.13
Power	50 kwh/T acid	$0.015/kwh	58.75
Total Utilities			786.92
M_gO	8.62 $T/10^3T$ acid	$120/T	81.03
Coke	6.06 $T/10^3T$ acid	$ 30/T	14.24
Catalyst	3.78 $gal/10^3T$ acid	$6.62/gal	1.96

(1) As 100% H_2SO_4

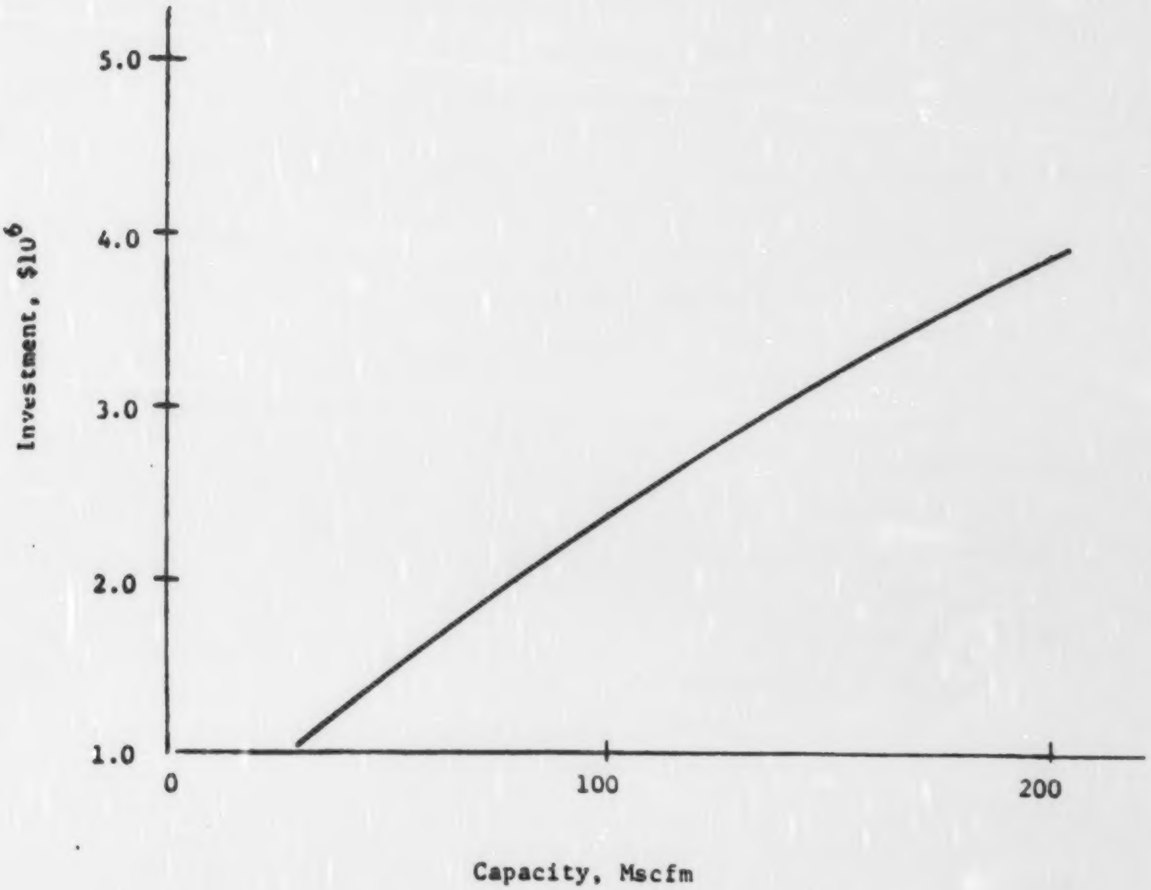


FIGURE L-9
MAGOX (CHEMICO) CAPITAL INVESTMENT
SCRUBBING SECTION

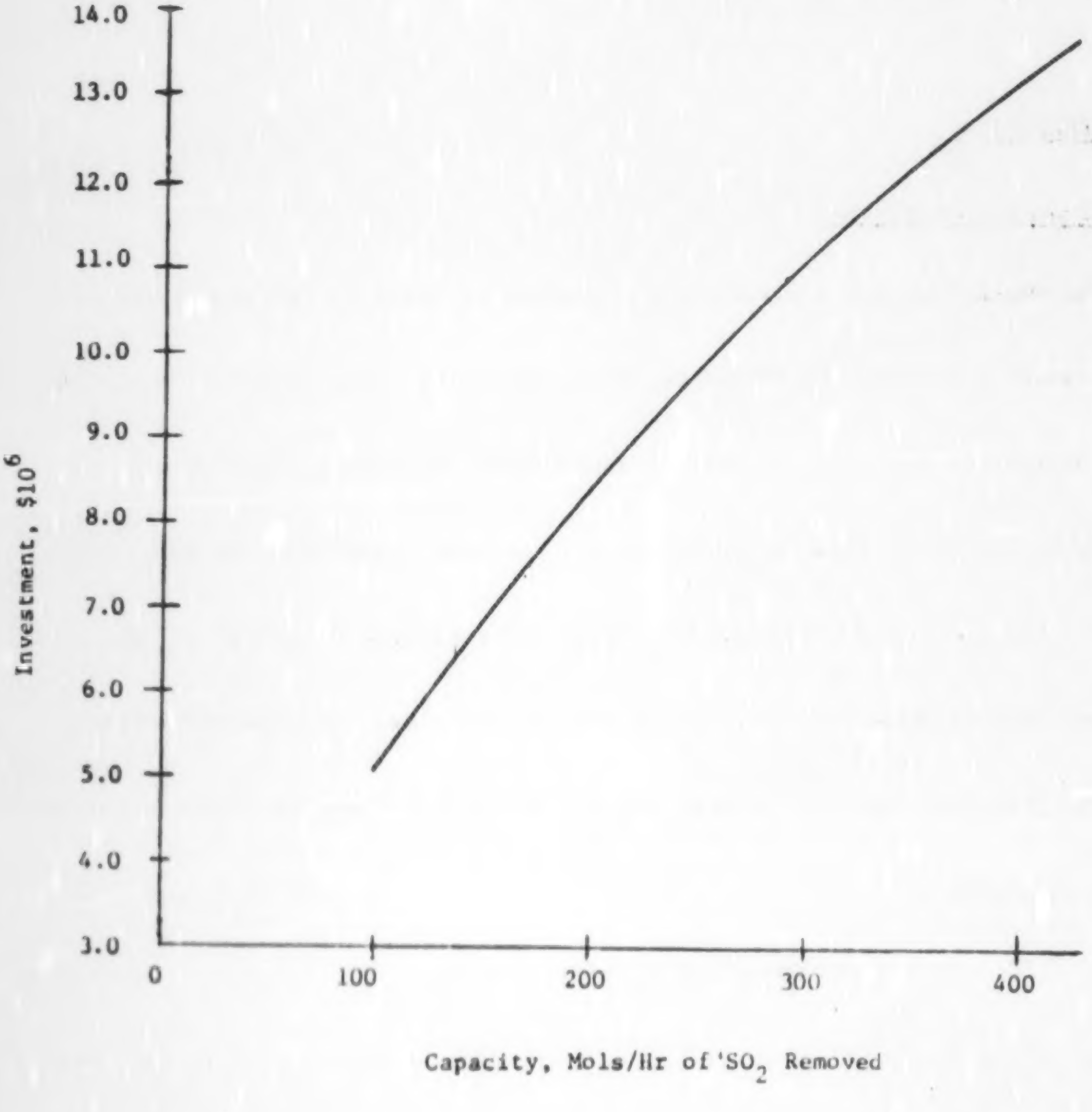


FIGURE L-10
MAGOX (CHEMICO) CAPITAL INVESTMENT
REGENERATION SECTION

5. SHELL/UOP

a. Process Description

The Shell flue gas desulfurization process is based on the reactivity of sulfur oxides with CuO. Fixed bed reactors containing a dry acceptor consisting of CuO-on-alumina are utilized for SO_2 adsorbtion. In situ regeneration by reducing gas is carried out at approximately the same temperature as SO_2 capture. The process was developed by Shell International Petroleum Mij at The Hague, Netherlands, and is licensed through Universal Oil Products Company.

The principal reaction between SO_2 in the flue gas and the copper activated alumina adsorbent is:

$$SO_2 + 1/2\ O_2 + CuO \rightarrow CuSO_4$$

The bulk of the accepted sulfur is released from the copper sulfate in the form of SO_2 upon regeneration with reducing agents such as hydrogen or carbon monoxide. With hydrogen, the overall regeneration reaction is

$$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O.$$

In addition, any unreacted CuO is reduced to copper. Some Cu_2S is also formed.

During the initial stages of the acceptance period, the Cu produced during regeneration is oxidized according to the following reaction:

$$Cu + 1/2\ O_2 \rightarrow CuO.$$

In addition, the Cu_2S present in the regenerated acceptor will oxidize as follows:

$$Cu_2S + 2\text{-}1/2\ Cu_2 \rightarrow CuO + CuSO_4$$

which will give the acceptor a based load of sulfur.

These oxidation reactions are exothermic and cause a considerable amount of heat to be released during the initial stages of the acceptance period. This results in a temperature peak which quickly travels through the reactor. The temperature peak is held below a predetermined level to protect the acceptor against a gradual loss of activity resulting from the growth of copper crystallites and possibly recrystallization of the active alumina. This is achieved by careful choice of the acceptor's copper content and of the reactor operating conditions.

Some of the more important process requirements are:

- Reactor inlet temperature between 700-800°F.

- Oxidizing conditions at the emission point to reactivate the regenerated acceptor.

- A source of reducing gas, preferably hydrogen.

Process advantages include:

- No loss of thermal buoyancy in the treated gas.

- No waste products since the process is dry and SO_2 is recovered as elemental sulfur.

- Minimum process water requirements.

- Low operating labor.

A simplified process flow schematic is shown in Figure L-11. Flue gas from the boiler passes through a rotary-type heat exchanger and is preheated by exchange with treated gas from the on-stream acceptor. Additional preheat is supplied by an indirectly fired trim heater, which supplies about 60°F of preheat. The 750°F flue gas passes vertically upwards through the on-stream reactor. The reactor internals are of "parallel passage" design with the acceptor contained between layers of wire gauze with spaces provided between packages for flue gas flow. With this design, the flue gas flows along the surface of the acceptor

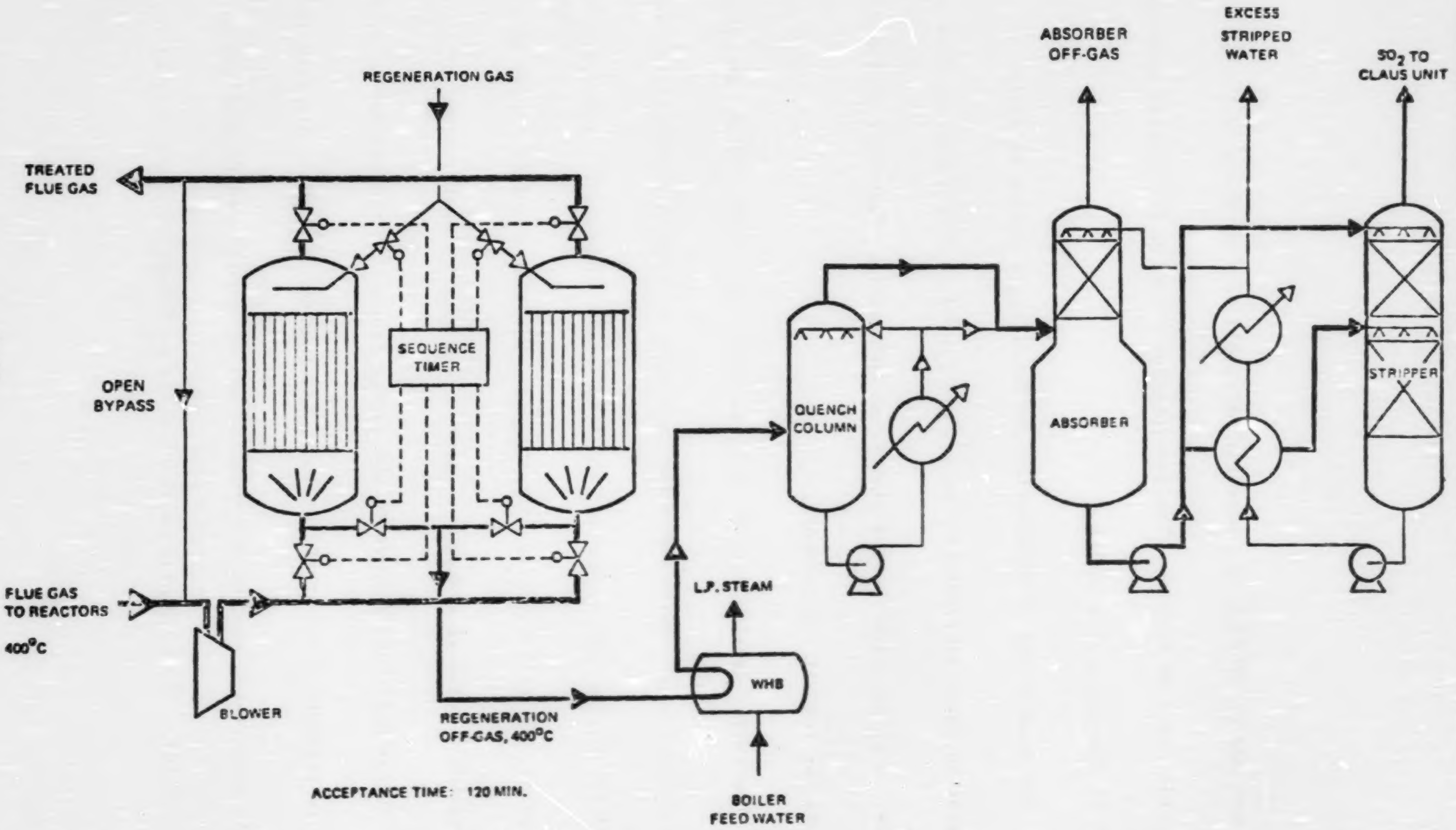

REGENERATION GAS
TREATED FLUE GAS
OPEN BYPASS
SEQUENCE TIMER
FLUE GAS TO REACTORS 400°C
BLOWER
REGENERATION OFF-GAS, 400°C
ACCEPTANCE TIME: 120 MIN.
L.P. STEAM
WHB
BOILER FEED WATER
QUENCH COLUMN
ABSORBER OFF-GAS
ABSORBER
EXCESS STRIPPED WATER
SO2 TO CLAUS UNIT
STRIPPER

Figure L-11 Simplified process flow scheme of SFGD packages and not through the acceptor material; thus plugging of the acceptor by particulates is avoided. The gas residence time in the thermal passages is about one-third of a second.

For convenience in fabrication and handling, a number of layers of acceptor packages appropriately spaced are placed in a container to form a module. A given reactor contains several such modules depending on the throughput and SO_2 removal requirements.

The sulfur dioxide will be accepted until the loading of the acceptor has proceeded to the point where the remaining unloaded acceptor at the exit end of the reactor has become too small to insure complete SO_2 removal. At this point, SO_2 will start to slip through. When the cumulative slip of the SO_2 reaches a value of 10% of the inlet concentration, purification of the gas input stream will have dropped to 90% and generation of the reactor is required.

To regenerate the spent reactors, the flue gas stream is diverted to the freshly regenerated reactor and reducing gas is simultaneously admitted to the spent reactor. The switching of reactors is performed automatically with the use of sequential control valves, thereby reducing labor requirements. The regeneration off-gas consisting of SO_2, H_2O and diluent leaves the reactor and passes through a waste heat boiler before entering the quench column. In the quench column, the regeneration gas is cooled and saturated before entering the SO_2 adsorber. In the adsorber, the regeneration gas is scrubbed with water to remove SO_2 before being vented, possibly to a fuel gas system. The water is removed from the adsorber and steam stripped to obtain concentrated SO_2. The stripped water is returned to the top of the adsorber. The concentrated SO_2 from the stripper can be sent to a contact unit to produce concentrated sulfuric acid or to a direct reduction process (Claus unit) to produce elemental sulfur.

In addition to the SO_2 product, the process has two other effluents. One of these is the stripped regeneration off-gases which are vented or can be returned to the boiler for recovery of combustion heat. The other stream is excess stripped water produced during regeneration which contains some dissolved SO_2. This purge stream can be sent to the boiler feed water treatment system to reduce raw water requirements.

b. Process Reliability

Shell began development work on the SFGD process about ten years ago. In order to prove the feasibility of the parallel passage reactor concept, a demonstration unit was erected in 1967 in the Shell refinery at Pernis, The Netherlands. The Pernis unit handled 600 scfm of flue gas containing 0.1-0.3 volume percent SO_2. The flue gas was isokinetically sampled from the main flue gas duct of a furnace fired with high sulfur fuel oil. A total of about 20,000 hours of testing was accomplished on this unit. The results of these tests convinced Shell that an acceptor based on a reinforced alumina support will have stable activity in excess of 8000 cycles or about 1-1/2 years of service life. The Pernis unit is still being used from time to time to test improved versions of the Cu-on-alumina acceptor.

After four years of successful operation of the Pernis unit, a decision was made to erect a full-scale unit at the Yokkaichi refinery of Showa Yokkaichi Sekiyu (SYS), Japan. The SFGD unit for SYS is designed to effect 90% desulfurization on 80,000 scfm (about 40 Mw), which is the combined flow of flue gas from an oil-fired boiler and tail gas from a Claus unit, the aggregate flow containing 2500 ppm SO_2. This unit was started up in the summer of 1973 and since then has only treated flue gas from the boiler since operation of the Claus unit has been discontinued for reasons not associated with the SFGD process. Based on limited information, the process availability appears to have been satisfactory.

As indicated in the process flow schematic, the reactor beds are automatically cycled from acceptor mode to regeneration mode by the use of sequential timers and motor-activated butterfly valves. To prevent leakage of reducing gas into the on-line reactor, special large, tight shut-off flue gas valves are required. In addition, pressure surges caused by reactor switching can influence boiler operation and are therefore undesirable. The effect of these surges is minimized by the use of an "open bypass" which continuously recycles a small quantity of treated flue gas. According to recent papers(14) given at the AIChE National Meeting in Tulsa, operations at SYS have confirmed the operability of the sequential switching and bypass system.

In using a fixed bed reactor for treatment of flue gas, one must consider the potential plugging problem associated with gas-borne fly ash. With the operation of the full-scale unit in Japan, this concern has been removed for oil-fired sources, which generate particulate in the range of 0.02-0.12 gr/scf of flue gas. Furthermore, UOP indicate that a conservative design figure for allowable inlet dust loading of 0.1 grains/scf.

In regard to applying SFGD to coal-fired boilers, a dummy parallel passage reactor was tested on a coal-fired boiler in The Netherlands to determine resistance to fouling by the unfiltered flue gas. No deposits were observed during a five-day duration run. Currently, tests on a coal-fired boiler are being performed at Tampa Electric with a single reactor containing a CuO acceptor. The unit is processing a slipstream from TECO's Big Bend Unit No. 1. The capacity of the unit is approximately three megawatts and is designed for a cycle time of 60 minutes to achieve 90% desulfurization. The objective of these tests is to demonstrate the applicability of this process to coal-fired sources.

c. Application to Refinery SO_x Control

The SFGD process is particularly suited for SO_x control in a refinery because:

- It produces a concentrated SO_2 gas stream which can be combined with the hydrogen sulfide from the amine unit and treated in a conventional Claus sulfur plant.
- The operation of the system is highly automated and labor requirements are minimal.
- The adsorbent accepts SO_2 and SO_3 equally well; consequently the system can be applied to the fluid catalytic cracker regenerator with no operating cost penalty.

As mentioned in an earlier section, the inlet temperature to the fixed bed reactor must be about 750°F. A refinery furnace or process heater designed to achieve reasonable thermal efficiency will normally have stack temperatures in the range of 400-500°F. Consequently, in a retrofit situation, preheating of the gas is required before entering the reactors. In order to minimize fuel requirements, a rotating heat exchanger (Ljunjstrom) is utilized to recover the heat in the treated gas by exchange with the incoming gas. A typical example of this arrangement is shown in Figure L-12. According to UOP(15), the inclusion of a heat exchange system increases the overall capital investment by about 15%.

In the refinery situation, supplying the required reducing gas is less difficult than for a utility boiler since hydrogen must be produced for other units within the refinery. Consequently, the cost of obtaining hydrogen can be assumed as incremental capital and operating cost for a reforming unit.

A refinery complex contains an atmospheric crude separation unit, numerous processing units and usually a central boiler facility, all of which contain stationary emission sources. In applying the Shell process to a typical refinery layout, it was assumed that several fixed bed acceptors would be located within the refinery to accept flue gas from several nearby emission sources. The alternative of placing individual reactors on each source becomes prohibitively expensive since the major capital investment for this process is in the acceptor section. The reducing gas is heated in a central location and distributed to the various acceptors. Regeneration gas from the acceptors is treated in a central absorption/stripping unit and the concentrated sulfur dioxide stream

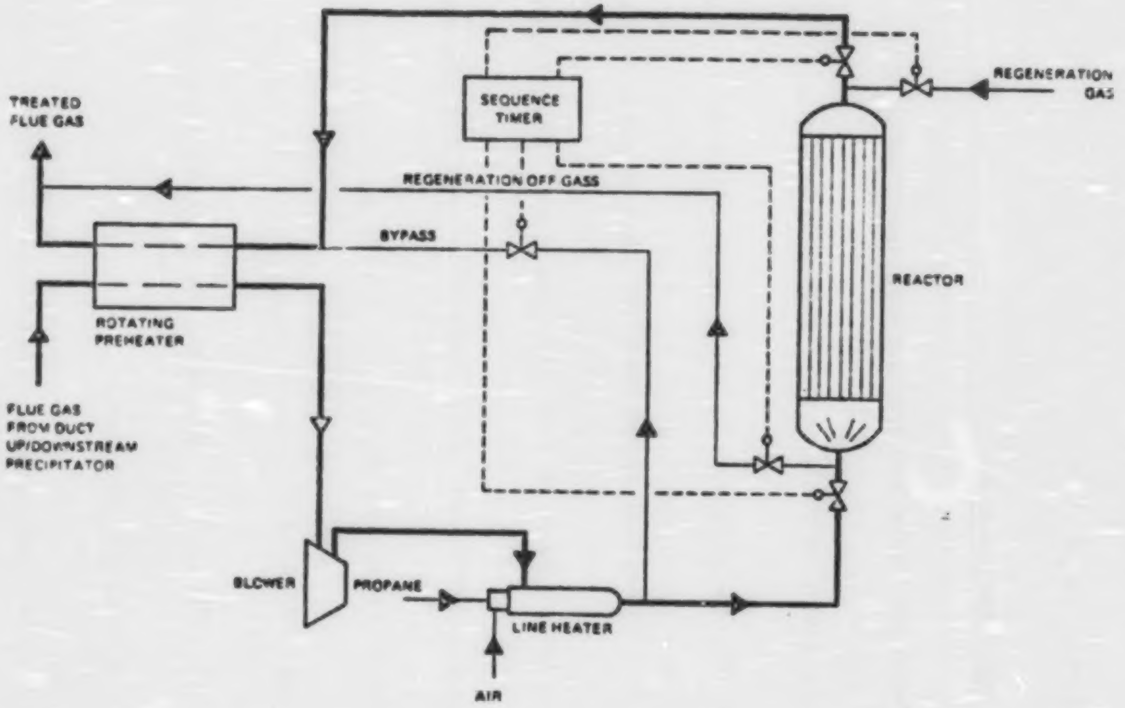

TREATED FLUE GAS
SEQUENCE TIMER
REGENERATION GAS
REGENERATION OFF GASS
BYPASS
REACTOR
ROTATING PREHEATER
FLUE GAS FROM DUCT UP/DOWNSTREAM PRECIPITATOR
BLOWER
PROPANE
LINE HEATER
AIR

FIGURE L-12

SIMPLIFIED FLOW SCHEME OF SFGD DEMONSTRATION UNIT
FOR COAL FIRED UTILITY BOILER AT TAMPA ELECTRIC, FLORIDA

leaving the stripper is combined with the sour gas stream from the amine unit and processed in a conventional Claus unit.

d. Capital and Operating Requirements

Typical capital investment costs for the acceptor section of a Shell flue gas desulfurization unit are shown in Table L-11. The estimated costs are based on a treatment capacity of 200,000 scfm and an SO_2 removal efficiency of 90%. Because the major portion of the capital investment for this process is associated with the acceptor system, a larger basic module size was used. This approach results in a savings in capital investment which more than pays for the additional duct work necessary for collection of the flue gas. Furthermore, the cost of a rotary-type heat exchanger was included to reduce annual fuel consumption required for preheat of the flue gases. Figure L-13 presents the capital costs for systems ranging in size from 30,000 scfm to 200,000 scfm.

Generalized costs for an SO_2 recovery and reduction unit were developed and are shown in Table L-12. The cost estimates are based on an SO_2 recovery rate of 200 lb-mols/hr or approximately 25,000 tons of elemental sulfur per year. In

TABLE L-11

CAPITAL & OPERATING COST ESTIMATE

Shell Flue Gas Desulfurization Acceptor System

Basis: 200,000 scfm
90% SO_2 Removal

Capital Investment ($M)	
Acceptor Section (Incl. Ton and Stacks)	3100
Rotary Heat Exchanger (Pre-heater)	500
Gas Collection Ducts	200
Process Directs	3800
Other Directs @ 10%	380
Total Directs	4180
Engineering and Contractor's Fee @ 22%	920
Subtotal	5100
Owner's Indirects @ 15%	765
Total Investment	5865

Annual Operating Requirements

Category	Unit Cost	Quantity	Annual Cost($M)
Labor	--	--	--
Utilities			
Power	$0.015/kwh	9.6×10^6 kwh	144
Fuel	$1.25/$10^6$ BTU	114.4×10^9 BTU	143
Total Utilities	--	--	287
Catalyst and Chemicals	$29/Ton of S	25,600 Ton	715

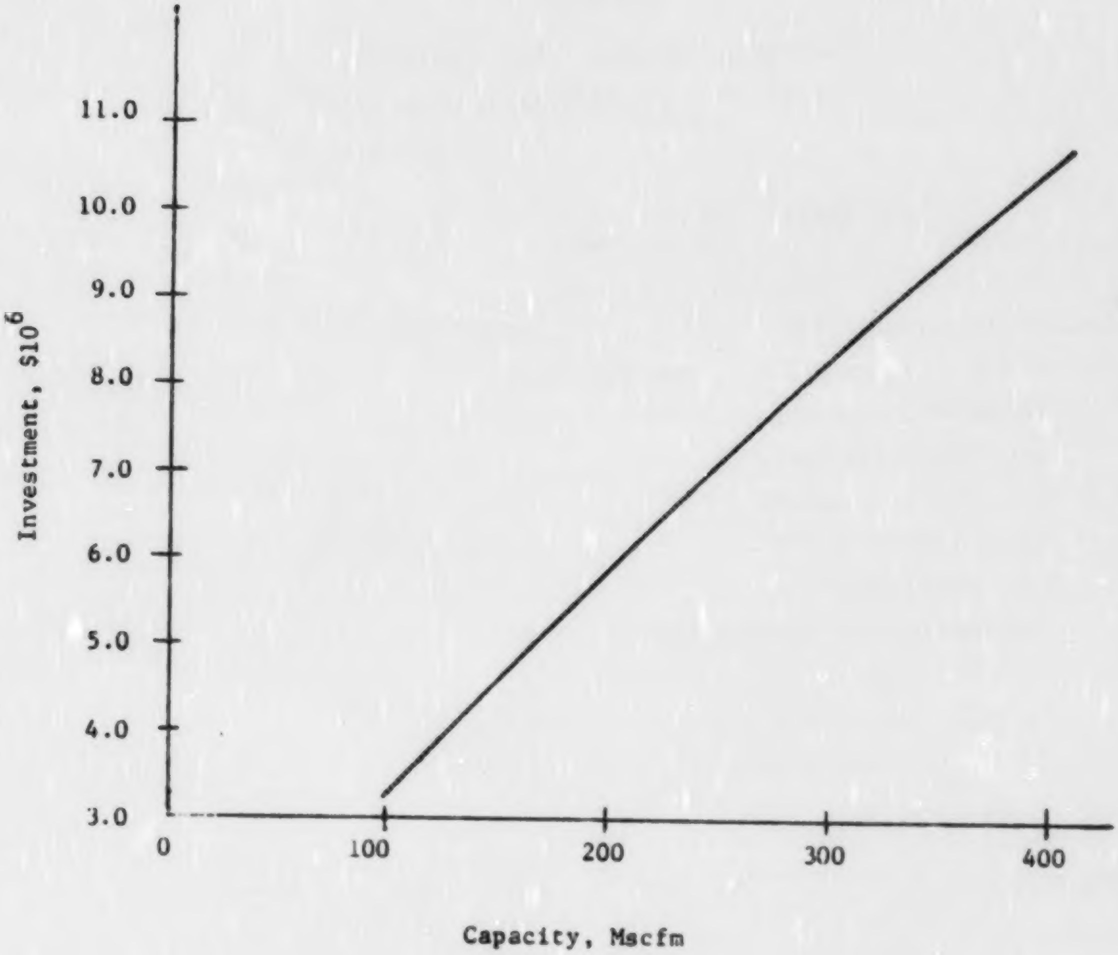


FIGURE L-13

SHELL/UOP, CAPITAL INVESTMENT
ACCEPTOR SECTION

TABLE L-12

CAPITAL & OPERATING COST ESTIMATE

Shell Flue Gas Desulfurization

Regeneration/Reduction Section

Basis: 200 # Moles/Hr of SO_2 Treated(1)

Capital Investment	
SO_2 Recovery Section	2790
Allowance for Interfacing	300
Process Directs	3090
Other Directs @ 10%	309
Total Directs	3399
Engineering and Contractors Fee @ 22%	748
Turn Key Claus Plant	1500
Subtotal	5647
Owner's Indirects @ 15%	848
Total Investment	6495

Annual Operating Requirements

Category	Unit Cost	Quantity	Annual Cost($M)
Labor	\$65,000/Shift Pos.(2)	1.5	98
H_2 Reducing Gas	\$0.70/M scf	1.75×10^5 M scf	1225
Utilities			
Power	\$0.015/kwh	5.35×10^6 kwh	80
BFW	\$0.80/M gal	7500 M gal	6
Cooling Water	\$0.05/M gal	8.0×10^4 M gal	4
Steam	\$1.25/M lbs	7.17×10^5 M lb	930
Total Utilities	--		
Claus Aunt Variable Cost	\$12.50/T of S	25,600	330

(1) 76.8 Tons of Sulfur/Day
(2) Full Coverage Operating Position

developing the capital investment costs, it was assumed that a central SO_2 recovery system would be located in the vicinity of the refinery sulfur plant (Claus unit). This requires a main regeneration gas header to and from the various acceptor reactors within the refinery complex. An allowance for interfacing the SO_2 recovery system with the acceptors has been included to account for this extra piping.

The concentrated SO_2 produced in the recovery section is combined with a portion of the refinery sour gas (H_2S) to obtain the proper stoichiometric ratio required for reduction to elemental sulfur in a conventional Claus unit. Included in the cost estimate is the investment required for a 77 ton/day sulfur plant including tail gas cleanup.

Estimated annual operating requirements for the recovery/reduction system are also presented in Table L-12. As can be seen, reducing gas and stripping steam represent the major utility items. Included with the operating requirements is a composite cost figure for the Claus plant variable costs shown in Figure L-14.

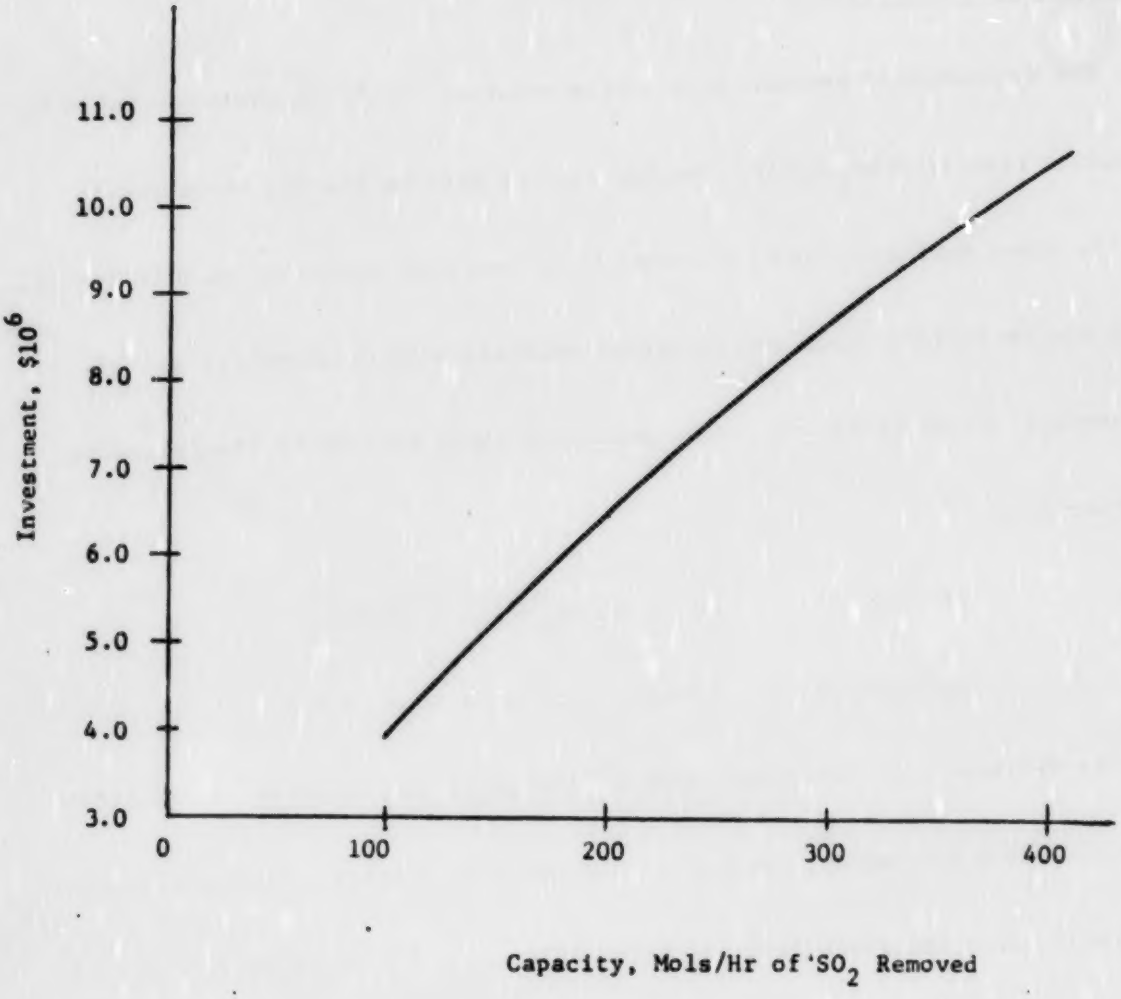


FIGURE L-14

SHELL/UOP, CAPITAL INVESTMENT
REGENERATION SECTION

6. WELLMAN-LORD

a. Process Description

The Wellman-Lord process is a sodium solution scrubbing system based upon a sulfite/bisulfite/SO_2 cycle. Sodium sulfite acts as the SO_2 absorbent and the spent sulfite liquor is thermally regenerated producing an SO_2-rich gas which can be further processed to either sulfuric acid or elemental sulfur. The overall system operation can be generally characterized by the following two reactions:

Absorption: $SO_2 + Na_2SO_3 + H_2O \rightarrow 2NaHSO_3$

Regeneration: $2NaHSO_3 \xrightarrow[\Delta]{} Na_2SO_3 + SO_2\uparrow + H_2O$

The Wellman-Lord technology takes advantage of the relative solubilities and equilibrium SO_2 vapor pressures of the two sodium salts. Sodium bisulfite has almost twice the solubility of sodium sulfite in the temperature range of the process. It is, therefore, possible to feed a solution to the absorption tower which is nearly saturated in sodium sulfite, since the solution composition is shifted in the direction of increasing solubility as SO_2 is absorbed.

This same solubility effect aids, in a reverse fashion, in the regeneration section. As SO_2 is evolved from the concentrated solution of sodium bisulfite, the sulfite salt is formed and rapidly reaches its solubility limit. The resultant precipitation of sodium sulfite helps to drive the regeneration reaction to the desired degree of completion with a minimum heat requirement.

Figure L-15 shows a flow schematic of the general process configuration for the Wellman-Lord SO_2 recovery system. In most applications there will be four processing steps in the recovery system--flue gas pretreatment, SO_2 absorption, absorbent regeneration, and purge treatment.

(1) Gas Pretreatment

The SO_2 absorber is usually designed to receive a saturated flue gas at 100-150°F which is free from particulate. Since the system operation is very sensitive to the buildup of contaminants, fly ash and other impurities in the gas must be removed or destroyed prior to contacting the SO_2 absorbent liquor. In the case of coal-fired boiler applications, pretreatment would involve fly ash removal using a high efficiency electrostatic precipitator followed by

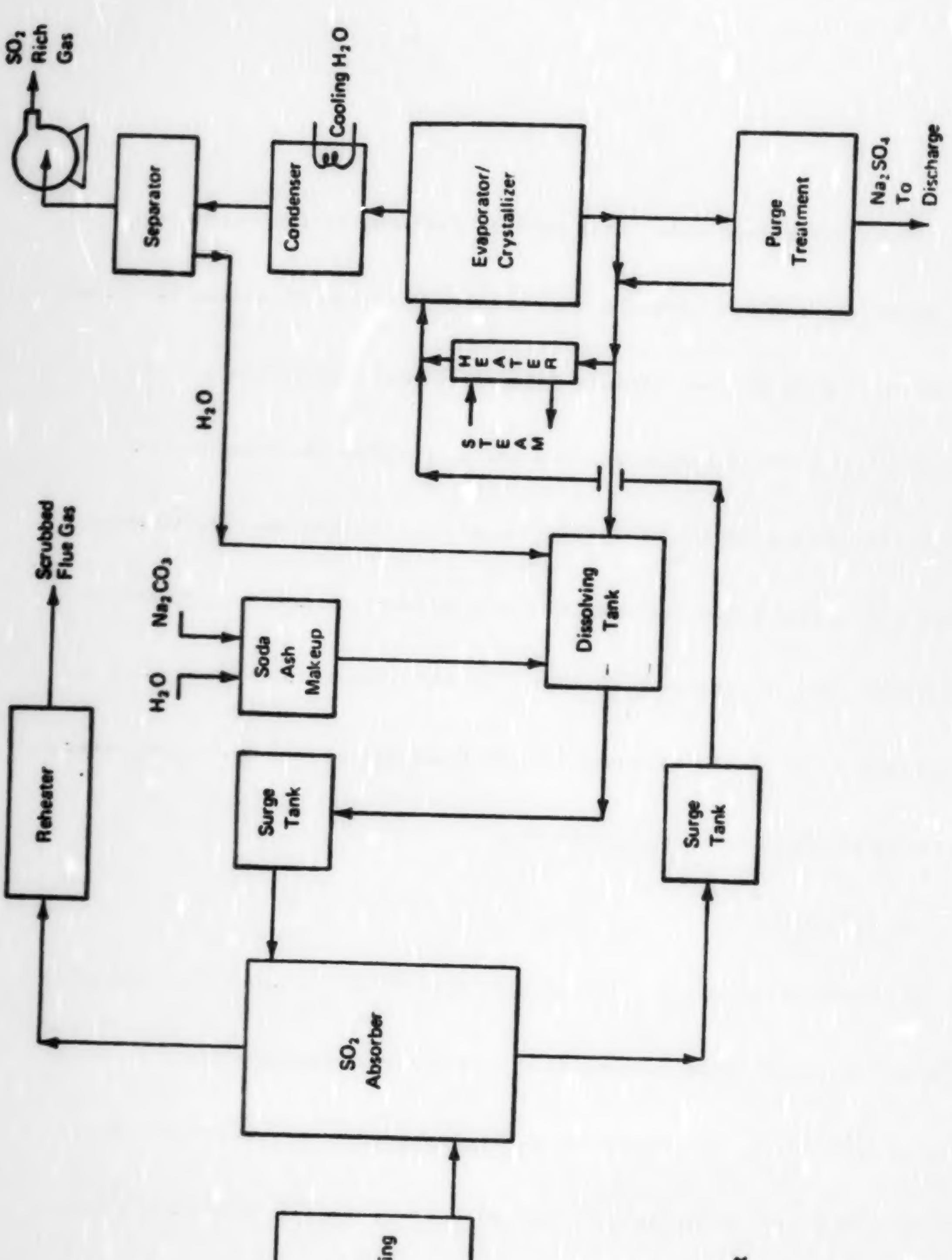

Low Particulate Flue Gas
H_2O Makeup
Gas Pre-Cooling
Purge to Treatment
SO_2 Absorber
Reheater
Scrubbed Flue Gas
Surge Tank
Surge Tank
H_2O
Na_2CO_3
Soda Ash Makeup
Dissolving Tank
H_2O
STEAM
HEATER
Evaporator/ Crystallizer
Purge Treatment
Na_2SO_4 To Discharge
Condenser
Cooling H_2O
Separator
SO_2 Rich Gas

FIGURE L-15 SCHEMATIC FLOWSHEET – WELLMAN-LORD PROCESS wet scrubbing to saturate the gas. The prescrubber not only cools the gas, but also ensures continued operation in the event of a failure of the electrostatic precipitator. In addition, it removes some small amount of fly ash (depending upon the particulate load and size distribution) as well as soluble fly ash components which could build up as impurities in the absorbent solution.

The wet scrubber most commonly specified is a low pressure drop venturi ($\Delta P = 4-6'' H_2O$). No absorbent is added to this cooling system, only fresh makeup water at a rate equivalent to the rate of water evaporation plus the liquor bleed rate. This liquor bleed, which is usually small (on the order of 50 gpm for a 400 MW system) can be returned to the prescrubber if it is relatively free from impurities. Otherwise, it must be neutralized (the bleed liquor pH is 1.5-2.0) and treated to meet local water pollution codes if it is to be discharged. The bleed rate is determined by the level of insoluble solids (<5 wt %) and the calcium concentration (to prevent $CaSO_4$ precipitation).

For applications to Claus plant tail gases it may be necessary to incinerate the gas to destroy H_2S, COS and CS_2 prior to SO_2 absorption. This incineration step is always followed by gas cooling via saturation with water. In other applications, such as oil-fired boilers, flue gas pretreatment may not be required because the flue gases are relatively "clean". However, provisions are usually made for gas prescrubbing prior to SO_2 absorption. This reduces the possibility of local precipitation of sulfite salts in the scrubber--a precipitation which could occur relatively easily if SO_2 inlet levels were to drop suddenly (this would decrease the amount of bisulfite formed, while water evaporation rates remain unchanged). In these applications (and even in some applications where pretreatment is used) in-line filters in the spent liquor stream are used to maintain insoluble contaminants at low levels.

(2) SO_2 Absorption

Following the prescrubber the gas passes through a de-entrainment separator and a mist eliminator to minimize the mixing of prescrubbing liquor with absorbent solution. The SO_2-rich gas is then contacted in a counter-current absorption tower with a concentrated solution of sodium sulfite and bisulfite (generally sodium concentrations are on the order of 6.0M Na^+). In the absorber, sodium sulfite is reacted with SO_2 to form sodium bisulfite:

$$SO_2 + Na_2SO_3 + H_2O \underset{\Delta}{\rightleftarrows} 2NaHSO_3$$

Sodium sulfate, which is nonregenerable, is also formed in the absorber both by the oxidation of sodium sulfite and be reaction of sodium sulfite and sulfur trioxide from the flue gas:

$$Na_2SO_2 + 1/2\ O_2 \rightarrow Na_2SO_4$$

$$2Na_2SO_3 + SO_3 + H_2O \rightarrow NaHSO_3 + Na_2SO_4$$

The resultant sulfate levels are controlled at a level of about 5 wt. % in the absorber feed stream by maintaining a continuous system purge.

The absorber itself is a multi-stage contacting device, such as a tray tower, with from three to six stages depending upon the required SO_2 removal. Liquor is recycled around each stage separately because the feed solution rate is generally not sufficient to adequately wet the trays (or packing). The largest absorption unit being will handle the flue gas from a 150-200 MW boiler.

Spent liquor bled from the recirculation leg on the bottom stage is discharged to a surge tank and then pumped to the regeneration system. Considerable surge capacity is provided for the absorbent feed as well as the spent liquor. This surge capacity ensures smooth operation during periods of fluctuating gas conditions and also provides for temporary shutdowns of the regeneration section.

(3) Absorbent Regeneration

The regeneration system basically consists of a conventional forced-circulation evaporator/crystallizer operated at a high internal recirculation rate. The evaporator can be designed to use low pressure steam (such as exhaust steam which might otherwise be discharged) although it is preferable to use high pressure steam; and the evaporator can also be either a single stage or multi-effect design (multi-effect is the usual type). A multi-effect configuration improves performance and can considerably reduce steam consumption. Multi-effect evaporators are used in the Japanese systems, however, in the United States, multi-effect evaporators require a considerably higher capital investment, so single effect evaporators are usually specified for small systems (less than about 150 MW). The largest evaporator now being designed is about 20 feet in diameter.

The primary reaction occurring in the regeneration system involves the conversion of sodium bisulfite to gaseous SO_2 and sodium sulfite, some of which precipitates as crystalline sodium sulfite:

$$2NaHSO_3 \rightarrow Na_2SO_3\downarrow + SO_2\uparrow + H_2O$$

There are, however, side reactions which produce byproducts which contaminate the liquor. These byproducts consist primarily of sodium sulfate, but also include some thiosulfate, thionate, and possibly elemental sulfur. It is not known whether the exact mechanism and kinetics of these side reactions is well defined, however, at least enough data has been compiled to guide the system design to reduce these undesirable byproducts.

The vapor leaving the evaporator is subjected to one or more stages of partial condensation to remove water. Existing plants are operating on both air and water-cooled condensers. To some extent, the degree of dewatering can be adjusted to provide any quality of SO_2 gas which is required for further processing (less than 10% water is easily achieved). The water obtained from this condensation is recycled to a dissolution tank and combined with the regenerated absorbent liquor.

(4) System Purge & Makeup

A continuous system purge must be maintained to prevent a buildup of contaminants in the system liquor. Possible contaminant sources include not only all byproduct formation in the system (Na_2SO_4, $Na_2S_2O_3$, S, etc.) but also all soluble and insoluble contaminants picked up from the flue gas and process makeup water. The most significant contaminant is sulfate. Since sulfate cannot be thermally regenerated it must be purged from the system in some type of bleed stream. This purge also removes other contaminants from the system, however the purge rate will normally be determined by the amount of sulfate formed. The effect of this purge requirement will be directly translated into process economics--in the capital investment for the purge treatment facility and the operating costs associated with the makeup sodium and purge disposal. Because these costs can be substantial and because process acceptability may be evaluated in terms of lost sodium value and waste disposability, a considerable process development effort has been expended in determining the various factors governing the rates of sulfate formation and developing methods for minimizing both sulfate formation and the attendant sodium losses in the required purge.

The majority of the sulfate is formed in the absorption system both from sulfite oxidation and from absorption of sulfur trioxide. To some degree the high solution concentrations used in the Wellman-Lord system tend to reduce the sulfite oxidation because oxygen solubility is greatly reduced at high solution concentrations. However, work is being directed toward further reducing oxidation by introducing anti-oxidants into the system.

In systems now in operation the purge stream is taken from the regenerated absorbent solution. This liquor must be treated to meet local water pollution codes which as a minimum would include neutralization and oxidation to eliminate COD. An alternative purge treatment approach which is under development involves crystallizing Glauber's salt from the regenerated liquor followed by filtration and drying to produce a relatively pure solid sulfate byproduct (about 90% Na_2SO_4 by weight). This purge treatment system will be incorporated into the NIPSCO installation that is now under construction.

In most cases the purge has amounted to about 10% of the absorbed sulfur value (the reported range in operating plants has been 5-20%), but it is expected that anti-oxidants will reduce this by 50%. Caustic or soda ash must be added to replace the sodium value lost in this purge stream. A purge containing 10% of the absorbed sulfur value tied up as sodium salts corresponds to a sodium makeup equivalent to 0.25 lbs. NaOH or 0.33 lbs. $NaCO_3$ per lb. S absorbed (0.14 lbs. NaOH or 0.18 lbs. Na_2CO_3 per lb. SO_2 recovered).

A research effort is currently being funded to develop a more sophisticated purge treatment (sulfate regeneration) system; however, no commercial approaches have yet been demonstrated. Most of this work is being conducted in Sweden.

As previously described, they are now treating the absorption system purge to produce solid, relatively pure sodium sulfate. The treatment and disposal of the purge from the gas prescrubbers has depended upon the particular type of application and system site. Where it cannot be mixed with sluiced ash, it must be neutralized and clarified and disposed of in an environmentally acceptable manner. Depending upon local water pollution codes, this could involve to be a troublesome and expensive treatment process.

b. Process Reliability

The Wellman-Lord process is a fully developed SO_2 control system for certain applications as evidenced by the number of commercial installations now in operation in the United States and Japan. In addition to the installations currently operating, shown in Table L-13, there are about a dozen systems either in the planning stage or presently under construction. To date most applications of this technology have been to either sulfuric acid or Claus plants and almost all have involved the conversion of the recovered SO_2 to sulfuric acid. There are, however, two systems in operation on oil-fired boilers in Japan and a retrofit system at Northern Indiana Public Service Company's (NIPSCO) coal-fired power plant in Gary, Indiana is due to start up in mid-1975. In addition, Davy Power Gas has recently been awarded the contract for a system on the San Juan plant of Public Service of New Mexico. The NIPSCO system

TABLE L-13

CAPITAL & OPERATING COST ESTIMATES
WELLMAN-LORD SCRUBBING SYSTEM

Basis:
- 100,000 scfm flue gas
- 2,200 ppm SO_2
- 90% SO_2 removal
- 8000 operating hours/year

CAPITAL COST ESTIMATES (Installed)

	Cost ($M)
Scrubber System	$ 420
Fan, Ductwork, and Stack	270
Gas Reheat System	60
Process Direct Cost	750
Other Directs @ 10%	75
Total Directs	825
Engineering & Fees (@ 22% TIC)	182
Subtotal	1,007
Owner's Indirects @ 15%	151
Total Investment	$1,158

ANNUAL OPERATING COST ESTIMATES (Full Load)

	Usage	Unit Cost	Annual Cost ($M)
Labor			
Utilities			
Fuel	50×10^9 Btu	$1.25/MM Btu	63
Steam			
Process Water	80×10^6 gal	40¢/Mgal	32
Power	3.9×10^6 kwh	1.5¢/kwh	59
Total Utilities			154
Purge Treatment	20×10^6 gal	$1.00/Mgal	20

will incorporate an SO_2 reduction process developed by the Allied Chemical Company for the production of elemental sulfur from the recovered SO_2.

The reliability of the Davy Power Gas/Wellman-Lord technology is indicated by the operating records of the installation at the Chuba plant of the Japan Synthetic Rubber Company. This system has operated from June, 1971 to March, 1973 with an on-stream factor of 97%, and from May 9, 1972 to March 1, 1973 with an on-stream factor of 100%. These stream factors refer to the scrubber circuit operation. Over this period the system has also achieved better than 90% SO_2 removal and has converted the recovered SO_2 to high quality sulfuric acid.

The major concern with the process operation has been the removal of contaminants and purge treatment and disposal, see attached.

Although difficulties in the evaporator/crystallizer operation have been few, the scrubber system operation is still protected by the considerable surge capacity (up to four days). This capacity also allows for continued absorption system operation during periods of routine system maintenance and when in-line suspended solids filters (if used) are being cleaned or replaced. The surge capacity also allows for the installation of a reasonable regeneration system size. The scrubber system itself requires no special back-up provisions since this is well-developed and proven operation which does not involve slurry handling.

c. Applicability to Refinery SO_x Control

The Wellman-Lord technology has several features that make it an attractive system for application to refinery SO_x emissions control problems. Specifically, these process features include:

- high SO_2 removal capability (>95%);
- low-scale potential in the scrubber system;
- ability to separate the scrubber system operation from the regeneration section, which allows the use of a centrally located regeneration facility serving a number of different scrubbers;
- generation of an SO_2 rich gas as the recovery system product that can be further processed in a Claus-type facility at the refinery; and
- high scrubber system stream factors due to the inclusion of surge capacity to cover regeneration system down time.

The Wellman-Lord process, though, also suffers from a number of inherent system limitations and potential operational disadvantages some of which may prove to be significant in a refinery application. These are:

- sensitivity of the system operation to the buildup of contaminants--this could be a major problem in scrubbing off-gases containing high SO_3 and O_2 levels or particulate loadings, conditions which may exist in the fluid cat-cracker flue gas (the Wellman-Lord system may not be a viable scrubber system for the fluid cat-cracker due to the potentially high SO_3 levels).

- the high saturation temperature of the recirculating liquid may require heat tracing of lines, particularly where the regeneration system is located some distance from the scrubbers;

- the scrubbed gas must be reheated to prevent plume formation;

  and

- the necessity of handling and disposing of a sizeable waste stream of soluble salts.

In general the Wellman-Lord process appears to be an appropriate control system for application to almost all refinery SO_2 emissions control problems. However, consideration must be given to the manner in which the system purge would be handled. There is also some uncertainty, regarding the suitability of the process for scrubbing the off-gases from the fluid cat-cracker units. The applicability to these units must be evaluated on an individual basis taking into consideration specific flue gas conditions, particularly SO_3 concentrations and particulate removal provisions.

d. Capital and Operating Requirements

(1) Scrubber System

Generalized costs have been estimated for installing and operating an SO_2 absorption system utilizing the Wellman-Lord technology. Table L-14 lists the capital investment (on an installed module basis) and operating cost breakdown for a unit treating 100,000 scfm of flue gas. Figure L-16 provides correlations of capital costs for systems ranging in size from 30,000 scfm to 200,000 scfm.

In developing these costs it has been assumed that gas precooling would be accomplished in a separate tower and that the unit is designed for 90% SO_2 removal (there is only a slight cost advantage for lower removal efficiencies). It has also been assumed that the scrubber system would be erected at ground

TABLE L-14

CAPITAL & OPERATING COST ESTIMATES

WELLMAN-LORD REGENERATION SYSTEM

Basis: • 200 # mols SO_2 treated/hour

CAPITAL COST ESTIMATES (installed)

	Cost ($M)
Evaporator/Crystallizer Section	$ 5,700
Purge Treatment System	1,300
Allowance for Scrubber/Regeneration Interfacing	800
Process Direct Cost	$ 7,800
Other Directs @ 10%	780
Total Directs	$ 8,580
Engineering and Fees (@ 22% TIC)	1,888
Turnkey Claus Plant	1,500
Subtotal	$ 11,968
Owner's Indirects @ 15%	1,795
Total Investment	$ 13,763

ANNUAL OPERATING COST ESTIMATES (at capacity)

	Usage	Unit Cost	Annual Cost ($M)
Regeneration System Variable Costs:			
Labor	3.25 Shift Pos.	$65,000/ShiftYr	211
Utilities			
Fuel	--		
Steam	$900x10^6$lbs steam	$1.25/Mlbs	1125
Water-Cooling	$6.5x10^9$gal(circ)	5¢/Mgal circ	325
Process	$1x10^6$gal	40¢/Mgal	∿ 0
Power	$9.0x10^6$kwh	1.5¢/kwh	135
Total Utilities			1585
Soda Ash	5400 tons	$50/ton	270
Waste Disposal (90%Na_2SO_4)	700 tons	Assumed = 0	0
Operating Supplies	--	--	20
Claus Plant Variable Costs			330

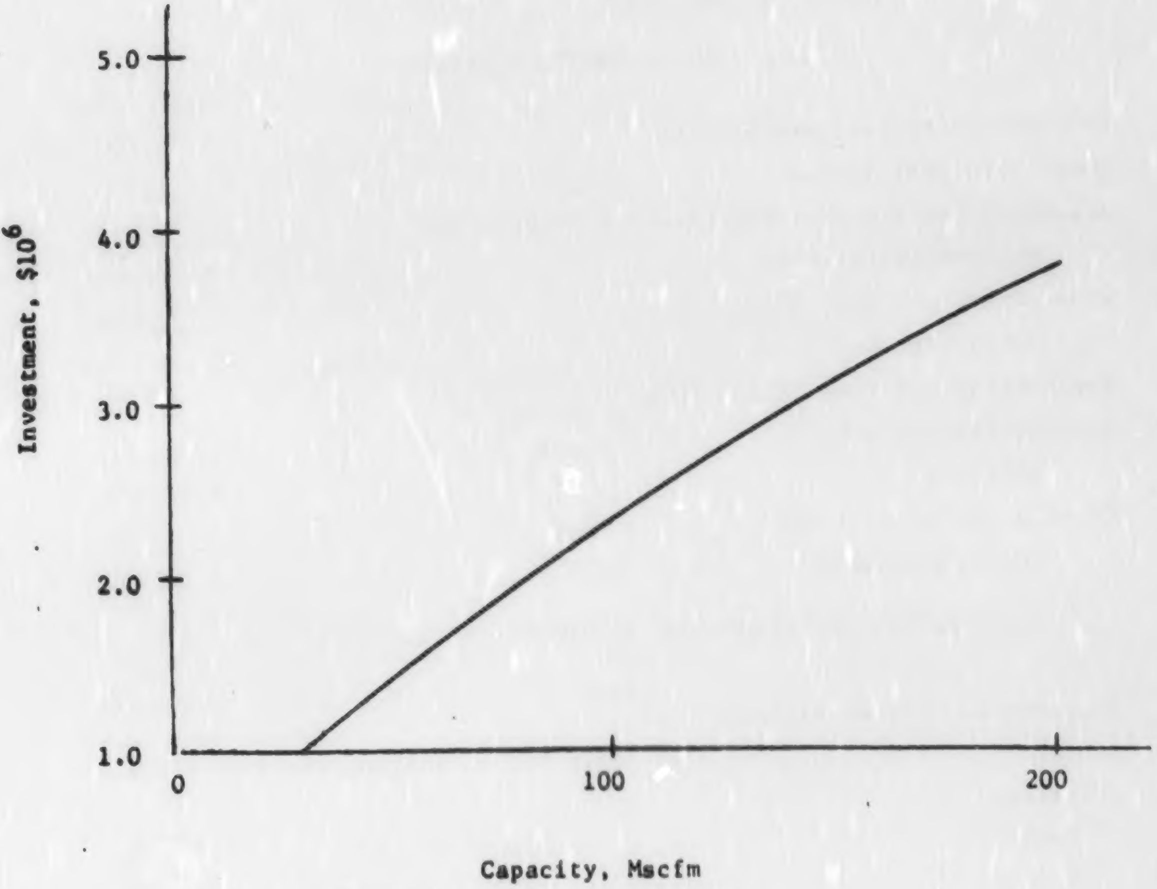


FIGURE L-16
DAVY POWER GAS, CAPITAL INVESTMENT
SCRUBBING SECTION

level and that there would be reasonable access to the scrubber site. Gas reheat would be by direct fired heaters. The purge treatment cost is for treating the acidic effluent from the precooler.

(2) Regeneration System

Cost estimates have been prepared for the thermal regeneration system on an installed module basis.

- The regeneration system would be a centrally located facility capable treating the effluent from all the refinery scrubbers. Table L-14 provides the costs for a system to recover 200 lb. moles/hr. SO_2. Figure L-17 shows capital and operating costs as a function of system size.

- It has been assumed that the purge treatment section would be similar to that for the NIPSCO installation and would produce a fairly pure Na_2SO_4. Allowance has also been provided for scrubber and regeneration system interfacing. This includes heated surge tanks with a one day storage based upon regeneration system capacity and a piping (heat traced) and control system.

  It has been assumed that the recovered SO_2 would be compressed and sent to a centralized Claus plant. The regeneration system costs include the SO_2-rich gas compressors, but not the duct work.

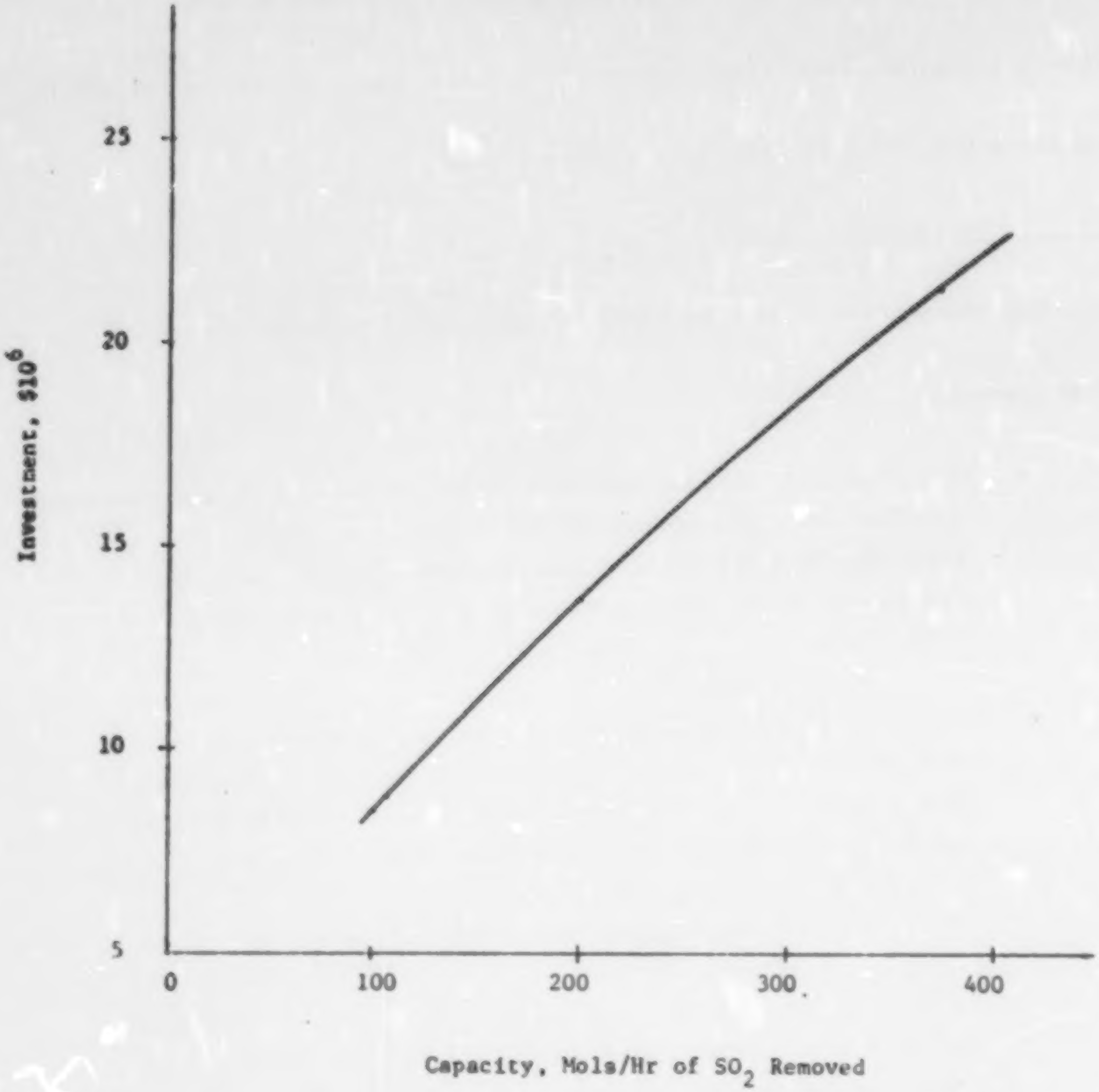


FIGURE L-17

DAVY POWER GAS, CAPITAL INVESTMENT

REGENERATION SECTION

C. OFF-LINE COMPARATIVE ECONOMIC ANALYSIS

As part of the process evaluation, annual operating costs for each of the five processes were developed. This economic evaluation was performed without the use of the refinery LP model and is hence referred to as an off-line analysis. However, to make the exercise meaningful, a basis was selected that would approximate the actual application to refinery sources. This was accomplished by determining capital investment requirements for each process based on six gas treatment systems of 100,000 scfm each and a central regeneration system with a capacity to handle the combined SO_2 removed by the six scrubbers or acceptors. The sulfur removal rate for this model is equivalent to 77 tons/day of elemental sulfur. This is nearly equivalent to the degree of removal required to attain 90% desulfurization of the flue gas from a 150 BPD refinery.

The results of this annual cost comparison are summarized in Table L-15. For those processes which generate concentrated SO_2 as the primary product, the investment required to convert SO_2 to a more convenient final product has been included.

TABLE L-13

FLUE GAS DESULFURIZATION PROCESSES
OFF-LINE COMPARATIVE ECONOMIC

Basis: 90% Sulfur removal from flue gas.
77 T/D Sulfur recovered.

	Chiyoda	Dual Alkali	MagOx	Shell/UOP	Davy Powergas Wellman Lord
Capital Investment, $M(1)					
Final Product	Gypsum	Calcium Sulfite	Acid	Sulfur	Sulfur
Gas Treatment Section	9,990	8,150	7,090	17,590	6,950
Regeneration Section	12,500	5,000	6,700	5,000	12,250
SO_2 Reduction	--	--	1,700	1,500	1,500
TOTAL INSTALLED COST (TIC)	22,490	13,150	15,490	24,090	20,700
$/Kw (Estimate)	83	49	57	89	77
$/Kw (Typical)(2)	--	26-47	36-66	(65)	40-68
Operating Cost, $M					
Utilities	2,163	798	1,777	1,880	2,510
Raw Material & Chemicals	1,642	1,660	110	1,900	290
Operating Labor	195	133	192	100	211
Waste Disposal	--	1,160	--	--	120
Maintenance (4% of TIC)	899	526	620	964	828
Fixed Charges (20% of TIC)	4,498	2,630	3,098	4,818	4,140
Claus Plant Operating Cost (3)	--	--	--	330	330
TOTAL ANNUAL COST	9,397	6,907	5,797	9,992	8,429

---

(1) Mid-1974.
(2) EPA presentation on Status of Flue Gas Desulfurization Technology - National Power Plant Hearings.
(3) Including Add-on Process.

For example, the cost of a sulfuric acid plant is included for the MagOx process and the investment for additional Claus plant capacity is added for the Shell/UOP and Davy Powergas processes.

The dollar per kilowatt equivalent of these estimates is shown in the table and are compared with typical values from other sources.(22) The estimates of this study are understandably in the high end of the range due to the complexity of interfacing this technology with typical refinery operations.

The annual operating costs for each process are also developed in Table L-15. Variable and semi-variable costs such as utilities, operating labor, catalyst and chemicals, and waste disposal are based on the individual process operating requirements presented in Section B. Maintenance and fixed charges are also included at 4% and 20% of total installed cost respectively. An important assumption made in determining the operating costs was that salable products such as gypsum, sulfuric acid, calcium sulfite and sulfur have zero market value and zero disposal costs. The zero product value assumption was made due to the prevailing uncertainties regarding the market for these products. However, it is conceivable that a firm might pay freight rates to obtain this material, thus alleviating a disposal problem for the producers. In the case of a throw-away process, disposal costs were set at $5/ton.

Based on this evaluation, the processes fell into two operating cost ranges. In the low range are the dual alkali and MagOx processes. Chiyoda, Davy Powergas and Shell occupy the high end of the spectrum. The most noticeable difference between the members of these two tiers is the degree of commercial application, which is undoubtedly reflected in the estimated annual cost. In general, annual costs tend to increase as processes near full commercialization.

The Wellman-Lord/Davy Powergas soda scrubbing process was selected to typify the economics associated with flue gas desulfurization in the refinery LP model. The reason for selecting this process is that it has been commercially applied in refineries in the United States and Japan and consequently the operating costs should be realistic. Furthermore, these costs are competitive with other processes exhibiting the same or similar degree of commercialization (i.e., Shell/UOP and Chiyoda). Finally, the process is compatible with existing refinery sulfur recovery processes which would make it attractive to refineries.

The one disadvantage with the process is that it may be unsuitable for application to the FCC catalyst regenerator due to possible high sulfur trioxide levels in the gas leaving the catalyst regenerator. Operating problems can be anticipated for processes based on the sodium sulfite/bisulfite system in this service. The main concern is a high rate of sulfate formation which would require a large system purge. This problem is diminished when a co-boiler is added to the system. This is due to a shift in the SO_2/SO_3 equilibrium concentrations at temperatures existing in the combustion zone. The SO_3 levels leaving the co-boiler are not much greater than those encountered in typical flue gases from fossil fuel combustion sources. .Consequently, the W-L process, with prescrubbers for particulate control, could be applied to the FCC.(12)

The MagOx process is an example of an emerging technology which appears environmentally and economically attractive. In the off-line evaluation, the overall economics were most attractive. However, these figures have more uncertainty than some of the other processes since the process has not operated continuously for any appreciable length of time. Hence, operating requirements are not as well defined. One of the major attractions of the process is that magnesium sulfate can be regenerated to recover the active magnesium. Hence, sulfate formation in the system is not a major concern. In addition, secondary emissions from the process are minimal. This process should evolve into a desirable candidate for flue gas desulfurization.

After completing this detailed assessment of FGD processes, it was learned that Exxon Research and Engineering has successfully applied caustic (NaOH) scrubbing technology to FCC regenerators for controlling particulate and SO_x emissions. The process flow schematic for this system shown in Figure L-18 is a standard, once-through arrangement frequently used with non-regenerable systems. A Jet Scrubber is employed to affect vapor-liquid contact and provide pumping capacity thereby eliminating the need for a fan. The scrubbing solution is the motivating fluid for the Jet Scrubber. A bleed stream is withdrawn from the circulating loop and delivered to a clarifier for removal of particulates. Solids and sulfur in the form of sodium sulfate

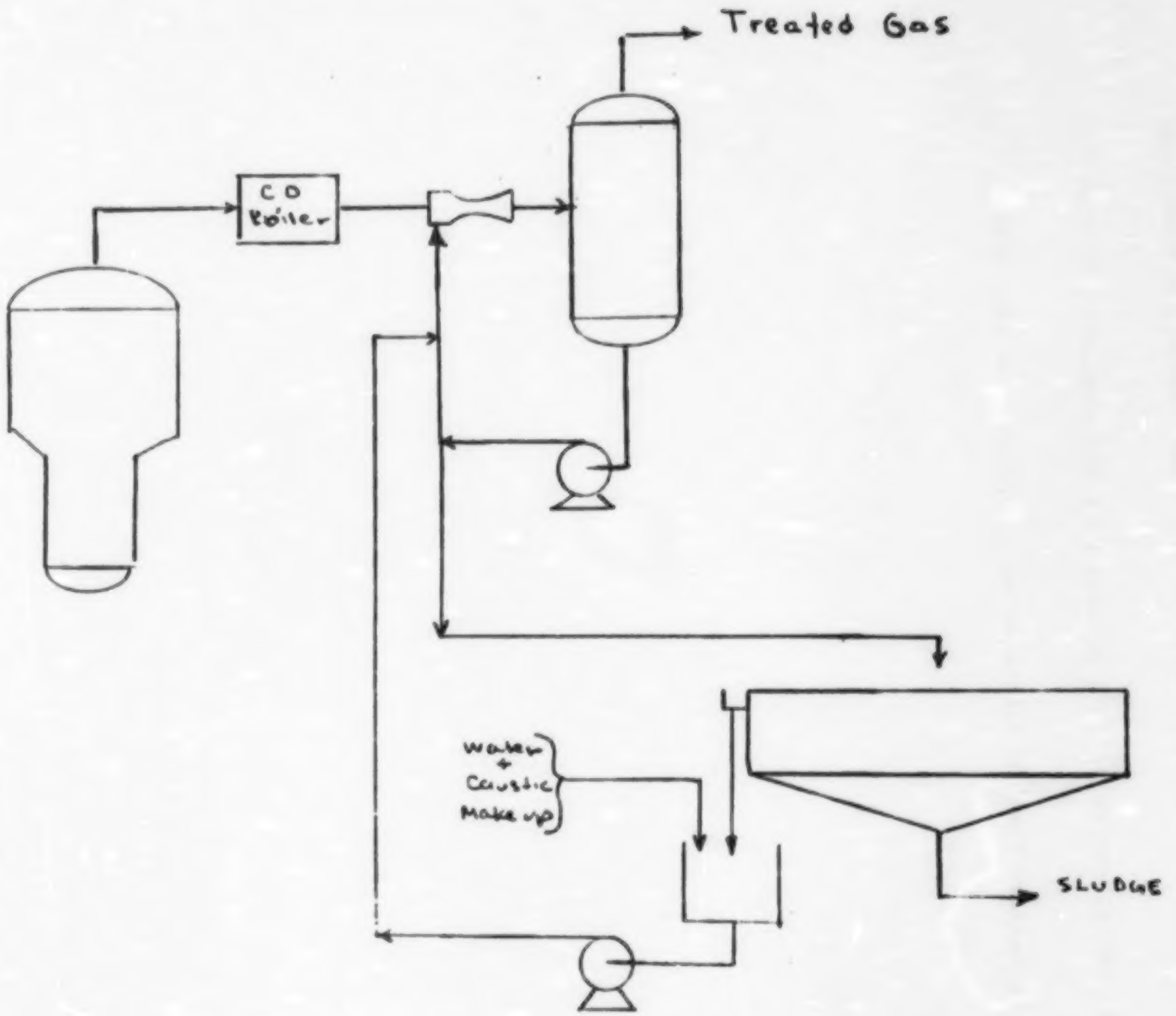

Treated Gas
CO
Boiler
Water
+
Caustic
Make up
SLUDGE

FIGURE L-18

TYPICAL FLOW DIAGRAM
EXXON FCC CAUSTIC SCRUBBING SYSTEM

are purged from the system in the thickened underflow. The clarified liquids with fresh caustic are returned to the scrubber.

Capital and operating requirements for the Exxon system are summarized in Table L-16 · The economics of this process were not available in time for comparison with Wellman-Lord to determine which process has the lowest overall cost.

TABLE L-16

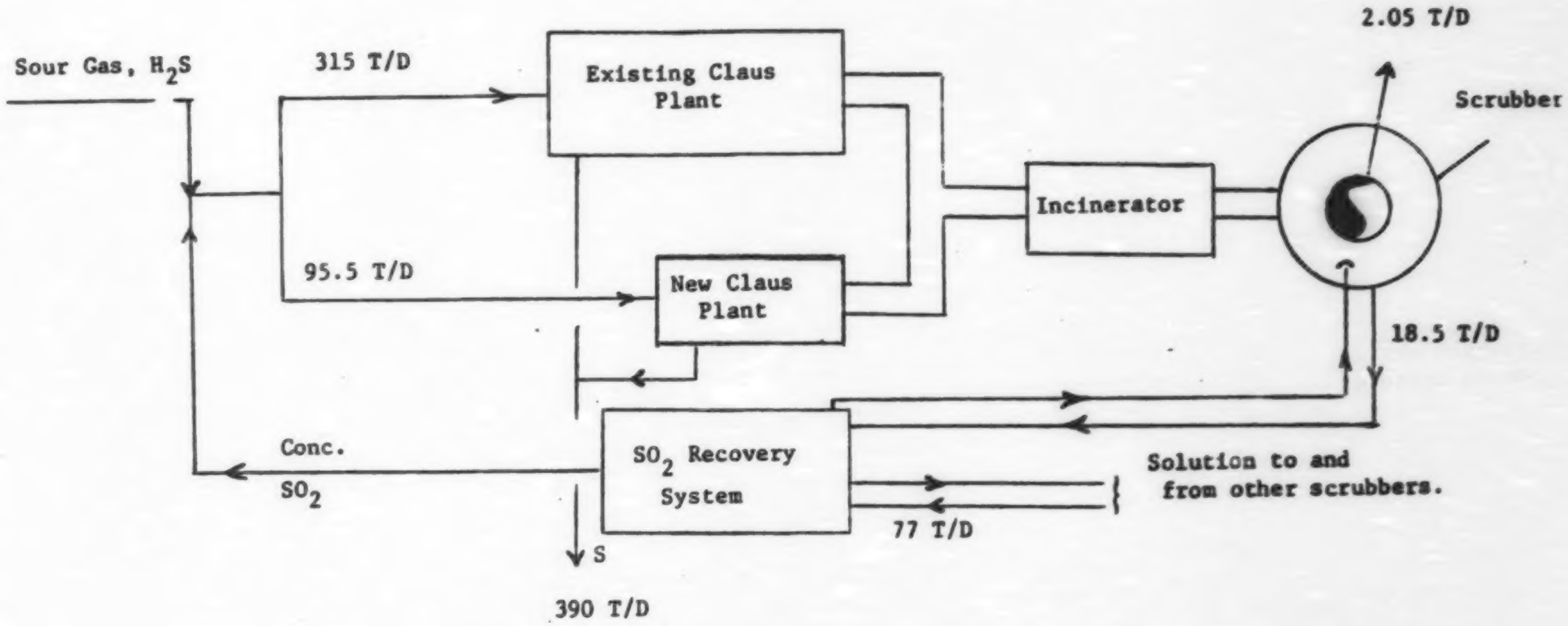
EXXON R AND E FCC SCRUBBING SYSTEM CAPITAL AND OPERATING REQUIREMENTS

Capacity, MScfm	55	125
Inlet Conditions		
Particulates, gr/Scf (dry)	0.1	0.1
SO_x, PPMV	1,000	1,000
Removal Efficiencies, %		
Particulates	78	78
SO_x	90	90
Capital Investment, $MM		
USGC; 1974	1.7	2.7
Utilities		
Makeup Water, gpm	85	200
Electric Power, kw	150	325
Fuelgas (10° reheat), MMBtu/hr	0.7	1.7
Caustic (26° Be), gpm	6.4	14.6
Sludge Generation Rate, gpm	0.75	1.7

D. CONTROL OF SULFUR PLANT EMISSIONS

1. ALTERNATIVES

The Claus sulfur plant tail gas from a typical three-stage converter contains between 7,000-12,000 ppmv of sulfur compounds.(22) Control of this source is necessary to achieve the required overall reduction in refinery emissions. In this study, two alternate routes were considered for treatment of the Claus plant tail gas. The first scheme, shown in Figure L-19, utilizes the same stack gas desulfurization technology ascribed to the other process units within the refinery. This approach involves the installation of a scrubber on the stack of the Claus plant incinerator. The spent absorbent is sent to the central SO_2 recovery system required for the other flue gas scrubbing units. The recovered sulfur value is recycled to the Claus plant along with the SO_2 removed from other sources. Scheme 2 assumes the installation of one of the available add-on processes such as the Beavon/Stretford or Pritchard Cleanair to the Claus unit to treat the

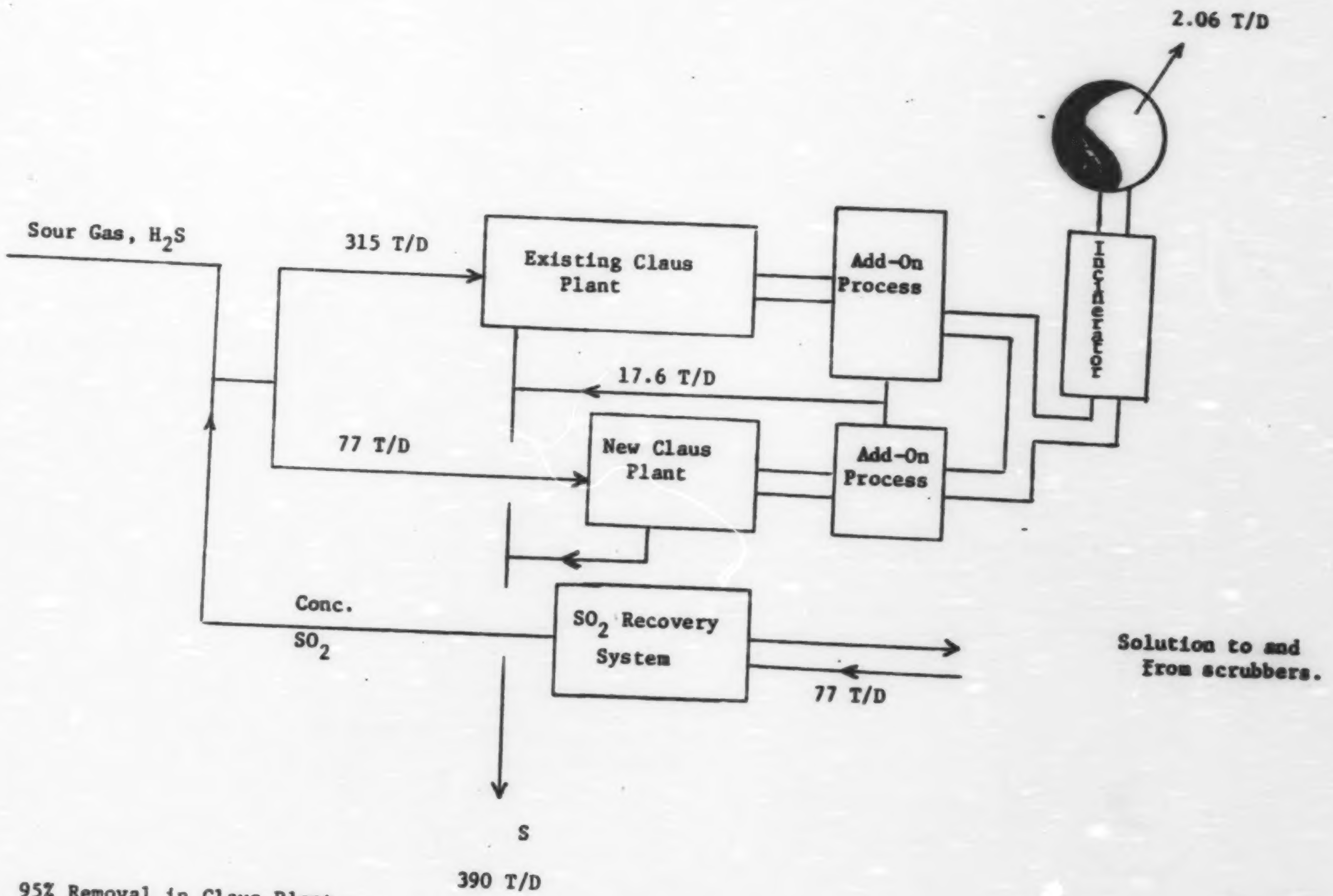

Sour Gas, H_2S
315 T/D
95.5 T/D
Existing Claus Plant
New Claus Plant
Incinerator
Scrubber
2.05 T/D
18.5 T/D
SO_2 Recovery System
Solution to and from other scrubbers.
77 T/D
Conc. SO_2
S
390 T/D

Basis: 95% Removal in Claus Plant
90% Removal in Scrubber
99.5% Removal of Fresh Feed

CLAUS TAIL GAS CLEANUP

SCHEME I

FIGURE L-19


2.06 T/D
Sour Gas, H_2S
315 T/D
Existing Claus Plant
Add-On Process
Incinerator
17.6 T/D
77 T/D
New Claus Plant
Add-On Process
Conc. SO_2
SO_2 Recovery System
Solution to and From scrubbers.
77 T/D
S
390 T/D

Basis: 95% Removal in Claus Plant
90% Removal in Add-On
99.5% Removal of Feed

CLAUS TAIL GAS CLEANUP

SCHEME II

tail gas. The gas leaving the add-on process is further incinerated to oxidize the remaining H_2S to sulfur dioxide.

2. ECONOMICS

The capital investment associated with Scheme I includes the cost of a scrubber system and incremental costs on the regeneration system and an incremental cost on the new sulfur plant to handle the recycle. The capital investment required for Scheme 2 is primarily the cost of the Claus add-on plant.

The economics of these two routes are summarized below:

	Scheme 1 Flue Gas Treatment	Scheme 2 Claus Add-On Process
Capital Investment, $M	2000	2100
Annual Operating Costs, $M	1180	650

The flue gas treatment costs for Scheme 1 are based on the figures developed in this report for the Wellman Lord process. The economics of the add-on process are based on available information in the published literature.

Based on this generalized evaluation, it appears that the Claus add-on process route has the more favorable economics. Consequently, we selected Scheme 2 for our sulfur control assessment study.

3. CLAUS TAIL-GAS-CLEANUP PROCESSES

A number of add-on processes are commercially available for treatment of Claus tail-gas. Some of the more advanced technologies include the following:

Process	Supplier
Beavon/Stretford	Ralph M. Parsons
CleanAir	Pritchard Co.
Sulfreen	IFP
SCOT	Shell

Of these, the Sulfreen process has the lowest sulfur removal capability, which is about 80-85%. The other processes all guarantee sulfur emissions less than 250 ppmv, or an overall recovery of better than 99.9 percent for the combination Claus plant/Tail-gas Cleanup System.

In our refinery SO_x control evaluation, we use the economics associated with the Beavon/Stretford process as typical of the cost for reducing sulfur emissions from Claus plants. The operating requirements for this process are shown in Table L-17.

TABLE L-17

BEAVON TAIL-GAS-CLEANUP PROCESS

TYPICAL INVESTMENT AND OPERATING REQUIREMENTS

Basis: 110 STPD Sulfur

	Per Short Ton of Sulfur
Capital Investment, \$MM (1st Quarter 1975)	\$2.75
Scale Factor	0.4
Operating Requirements	
Electric Power, kwh	284
Steam, Mlbs (50 psig)	0.14
Fuelgas, 10^6 Btu	4.36
BFW, Mgal	0.13
Labor, Oper/Shift	1/2

E. INTEGRATION OF SO_2 REMOVAL PROCESSES

This section describes in a general way how the selected process might be integrated into a typical refinery complex.

1. DAVY POWERGAS PROCESS

A conceputal refinery SO_x control system based on the Davy Powergas process is presented in Figure L-20. Basically, individual flue gas scrubbers are installed on the various refinery units including the FCC, and the spent adsorbent from each scrubber is sent to a central regeneration SO_2/recovery system with the regenerated adsorbent returning to the scrubbers. The recovered SO_2 stream containing about 95% SO_2 is sent to a Claus sulfur plant.

The concentrated sulfur dioxide from the central regeneration recovery system is combined with the sour gas from the amine unit to form the feed for the sulfur plant. A new sulfur plant is also required since the capacity of the existing unit is likely to be insufficient to handle all of the SO_2 recovered from the stationary sources. The combined tail-gas from both sulfur plants is subjected to a tail gas cleanup step, employing add-on process technology as discussed in Section D.

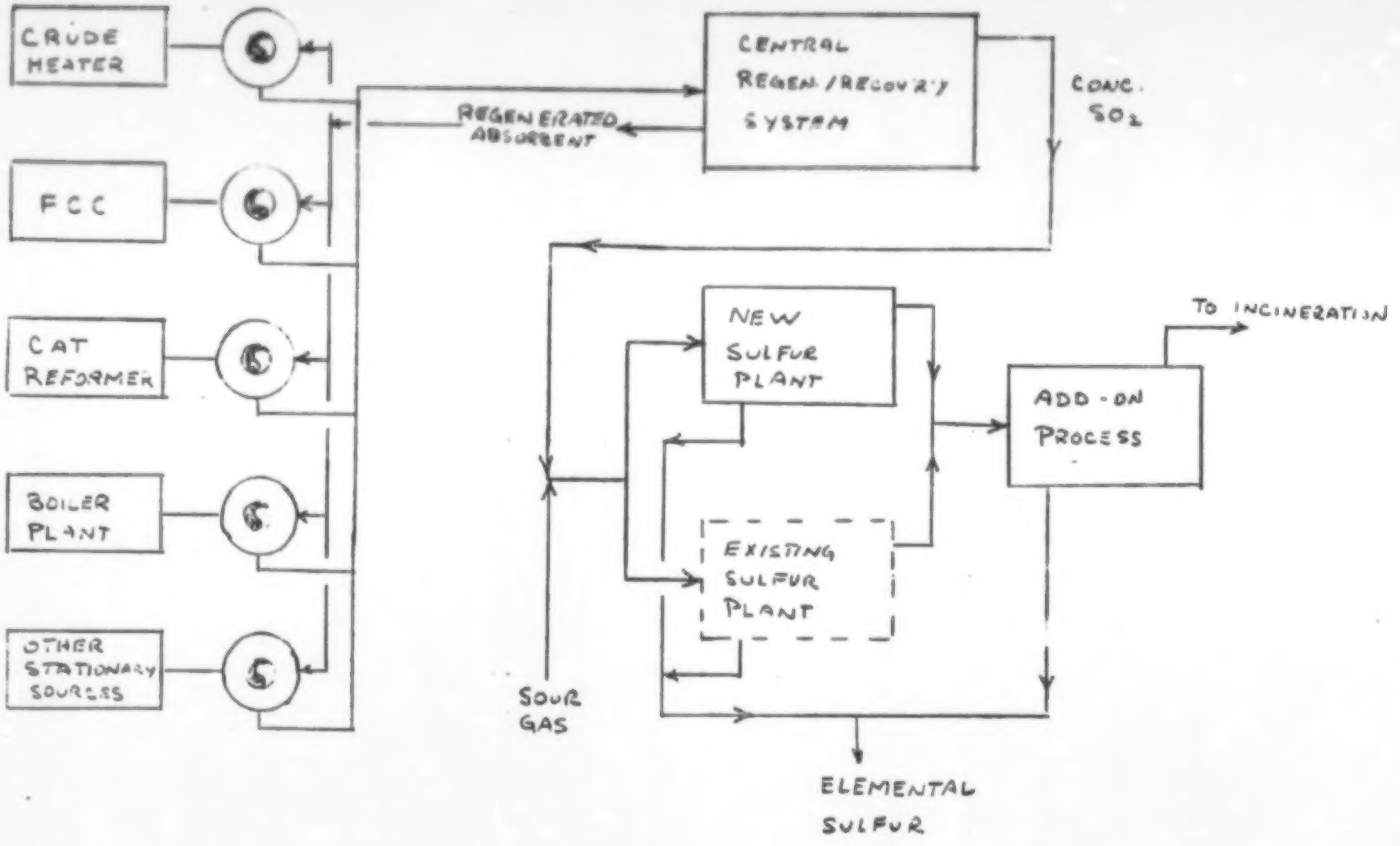

CRUDE HEATER
FCC
CAT REFORMER
BOILER PLANT
OTHER STATIONARY SOURCES
CENTRAL REGEN./RECOV'R'Y SYSTEM
REGENERATED ABSORBENT
CONC. SO₂
NEW SULFUR PLANT
EXISTING SULFUR PLANT
ADD-ON PROCESS
TO INCINERATION
SOUR GAS
ELEMENTAL SULFUR

FIGURE L-20

CONCEPTUAL REFINERY SO_x CONTROL SYSTEM

BASED ON WELLMAN-LORD PROCESS

2. PROCESS REQUIREMENTS

The capital investment/capacity relationships associated with refinery flue gas desulfurization are shown in Table L-18 for the process programmed into the refinery model. The investment costs are consistent with the first quarter 1975 basis used in the model. The investment cost equation for the regeneration/recovery section does not include the cost of the additional Claus unit. This was excluded since sulfur plant economics are handled separately in the refinery LP model.

Process operating requirements for the FGD processes are presented in Table L-19. The operating requirements are presented in terms of consumption per unit of treatment volume or weight of sulfur produced.

In applying flue gas scrubbing in a cost effective manner, scrubber sizing must be considered. For sources with a treatment volume of less than 30 Mscfm, it is more economical to combine the flue gas with that of a nearby source and take advantage of economies of scale. However, for large sources such as the atmospheric crude heater or the boiler facility, scrubber availability becomes a major consideration. Consequently, for emission sources in

TABLE L-18

FLUE GAS DESULFURIZATION PROCESS ECONOMICS
CAPITAL REQUIREMENTS

GAS TREATMENT

- Capital Investment (Mid 1975)

Process	Davy Powergas Wellman Lord
Capacity, Mscfm	100
Total system cost, \$M(1)	1275
Size factor	$[\text{Mscfm}/100]^{0.7}$

ABSORBENT REGENERATION/SO_2 RECOVERY

- Capital Investment (Mid-1975)
- Basis: 200 lb mols SO_2/hr treated(3)

Process	Davy Powergas Wellman Lord
Total system cost, \$M	15,125(2)
Size Factor	$\left(\frac{\text{lb mols } SO_2/\text{hr}}{200}\right)^{0.7}$

(1) Includes acid plant.
(2) Excludes Claus plant.
(3) 76.8 T/D of Sulfur.

TABLE L-19

REFINERY FLUE GAS DESULFURIZATION PROCESS OPERATING REQUIREMENTS

PROCESS:	Davy Powergas Wellman-Lord
Scrubber Section	
Electric Power, kwh/Mscf	0.0813
Fuel, Btu/Mscf	1040
Process Water, Mgal/Mscf	0.00170
Purge Treatment, $/Mscf	0.00042
Regeneration Section	
• Utilities	
Steam, Mlbs/T of S	35
Electric Power, kwh/T of S	352
Process Water, Mgal/T of S	0.039
Cooling Water, Mgal/T of S	254
• Raw Materials	
Soda Ash, $/T of S	10.50
• Waste Salts, T/T of S	0.27
• SO_2 to Claus Plant T/T of Product	2.0
• Sulfur Product	Conc. SO_2

excess of 100,000 scfm, two 60% scrubbers should be considered. In summary, recommended scrubber utilization is as follows:

Flue Gas Treatment Volume, Mscfm	Number of Scrubber Systems Required
<30	Combined with another small source
30-100	1
100-250	2

REFERENCES

1. National Public Hearings on Power Plant Compliance with Sulfur Oxide Air Pollution Regulations, U.S. Environmental Protection Agency, January 1974.

2. "An Introduction to Stack Gas Cleaning Technology," A.V. Slack, Technical Conference on Sulfur in Utility Fuels: The Growing Dilemma, October 25-26, 1972.

3. "Japan Tries Three Stack-Gas Desulfurization Routes," The Oil and Gas Journal, July 24, 1972, pp. 36-41.

4. New Flue Gas Desulfurization Process, Hideo Idemura, June 20, 1973.

5. The Chiyoda Thoroughbred 101 Flue Gas Desulfurization Process, Chiyoda Chemical Engineering & Construction Co., Ltd., 1973.

6. Summary Report on U.S. Requirements for Desulfurization, Battelle Columbus Laboratories, November 22, 1972.

7. "Conceptual Desig [illegible] Cost Study, Sulfur Oxide Removal from Power Plant Stack Gas, Magnesia Scrubbing--Regeneration: Production of Concentrated Sulfuric Acid," G. G. McGlamery, R. L. Torstrick, J. P. Simpson, J. F. Phillips, Jr all of TVA, Muscle Shoals, Alabama, 372 pages, Office of Research and Monitoring, U.S. EPA-R2-73-244, May 1973, Washington, D.C.

8. "Operational Performance of the Chemico Basis Magnesium Oxide Systems at the Boston Edison Company," Part I, George R. Koehler, Chemical Construction Corp., New York, N.Y., Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973, 25 pages.

9. "Operational Performance of the Chemico Magnesium Oxide System at the Boston Edison Company," Part II, Christopher P. Quigley, Boston Edison Company, Boston, Massachusetts, Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973, 12 pages.

10. "Some General Economic Considerations of Flue Gas Scrubbing for Utilities," John K. Burchard, Garty T. Rochelle, William R. Schofield, John O. Smith, Technical Conference on "Sulfur in Utility Fuels: The Growing Dilemma," sponsored by Electrical World, October 25-26, 1972, pp. 91-124.

11. "Control of Sulfur Oxides in Stack Gases; Magnesium Base SO_2 Recovery Process: A Prototype Installation at Boston Edison Company and Essex Chemical Company," I. S. Shah, C. P. Quigley, Symposium No. 39E, Seventieth National Meeting, Atlantic City, N.J., August 29-September 1, 1971, AIChE, New York, N.Y., 30 pages.

REFERENCES (Continued)

12. "Refinery Catalytic Cracker Regenerator SO_x Control," T. Ctvrtnicek, T. W. Hughes, C. M. Moscowitz, D. L. Zanders, Contract No. 68-02-1320, Task No. 1, Phase I, EPA Control Systems Laboratory, Chemical Processes Section, Research Triangle Park, North Carolina, September 1973, 161 pages.

13. Conversation with M. A. Maxwell, EPA Project Director for Boston Edison/EPA Mag-Ox Study.

14. "Shell Flue Gas Desulfurization Process Demonstration on Oil and Coal-Fired Boilers," AIChE National Meeting, Tulsa, Oklahoma, March 1974.

15. "New Tool Combats SO_2 Emissions," R. E. Conser, R. F. Anderson, *Oil and Gas Journal*, October 29, 1973.

16. "Dry Process for SO_2 Removal Due Test," *Oil and Gas Journal*, August 21, 1972.

17. *Control of Sulfur Oxide Pollution from Power Plants*, F. T. Princiotta and N. Kaplan, EPA Control Systems Divisions Report, October 1971.

18. *Economics of Flue Gas Desulfurization*, G. T. Rochelle, presented at the Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973.

19. Discussions with C. B. Earl, Davy Powergas, Lakeland, Florida.

20. Discussions with R. Christnan, Federal EPA Project Officer, Research Triangle Park, North Carolina.

21. *The Wellman-Lord SO_2 Recovery Process*, B. H. Potter and C. B. Earl, presented at 1973 Gas Conditioning Conference.

22. "Environment Needs Guide Refinery Sulfur Recovery," H.S. Bryant, *Oil and Gas Journal*, March 26, 1973.